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OPTIMIZATION AND DESIGN OF AN OIL ACTIVATED SLUDGE CONCENTRATION PROCESS

T. M. Rosenblatt

Esso Research and Engineering Company

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16. Abstract

Laboratory and pilot plant studies are described for a new Esso-Carver Greenfield process for the disposal of sewage sludge. An oil-assisted gravity separation of the majority of the water while heating is followed by multiple effect evaporation to dryness in an oil slurry, and incineration of the dry solids. Agreement between laboratory and pilot plant results was good, indicating no scale-up problems.

In the gravity separation, secondary sludges are concentrated from about 0.5% up to 5-10% solids. Solids capture of 98% or more is achieved by high shear oil-sludge contacting. Temperature dependent losses of solubilized organic carbon up to about 25% of the organic content of the feed are observed in the separated water from the oil concentration, and in the distillate from the evaporators. The process economics show an advantage of $13-32 a ton compared to the best known commercial technology: total costs are estimated at $21-39/ton of dry solids for a 189 ton/day plant processing a 50/50 mixture of primary + activated sludges to ash. A lower temperature gravity separation step could greatly reduce the economic penalty for a 25% recycle of solubilized secondary sludge, and yield an improvement of $1-12/ton of dry solids depending on plant size and sludge type. Other cost reductions in the thickening and settling steps could amount to $1-5/ton dry solids. (EPA)

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Program Element 1BB043

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Cincinnati, Ohio 45268

Prepared for

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ABSTRACT

This report describes laboratory and pilot plant studies and cost calculations for a new process for the disposal of sewage sludge. The process consists of an oil assisted gravity separation of the majority of the water, with heating, followed by multiple effect evaporation to dryness in an oil slurry and incineration of the dry solids.

In the gravity separation, secondary sludges were concentrated from about 0.5% up to 5-10% solids. Solids capture was $\geq$ 98% with high shear oil-sludge contacting. However, solubilized organic carbon losses were observed in the separated water from the oil concentration, and in the distillate from the evaporators. These losses were primarily temperature dependent and ranged up to about 25% of the organic content of the feed. The agreement of performance between laboratory and pilot plant results was good, indicating no scale-up problems. The process economics show an advantage of \$13-32 a ton compared to the best known commercial technology, for a 189 ton/day plant processing a 50/50 mixture of primary + activated sludges to ash. The total costs for the process are estimated at \$21-39/ton of dry solids for the 189 ton/day plant. These cost estimates include an economic penalty for a 25% recycle of solubilized secondary sludge. A lower temperature gravity separation step would greatly reduce the total solubilization loss and could yield a net economic improvement of \$1-12/ton of dry solids, depending on plant size and sludge type. Other possible cost reductions in the thickening and settling steps have been identified which could amount to \$1-5/ton dry solids.

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GLOSSARY OF ABBREVIATIONS

BOD_5 = Biological Oxygen Demand, as determined in a 5-day test

HLB = Hydrophilic/Lipophilic Balance, characteristic of a surface-active material.

HO = Heating Oil

LOPS = Low-Odor Paraffin Solvent (brand name)

MGD = Thousands of Gallons per Day

TC = Total Carbon, as determined on an aqueous sample by ASTM test D-2579-69 (APHA-138A).

TOC = Total Organic Carbon, same procedure.

TEI = Total Erected and Installed cost (Esso Engineering estimates). All costs are as of March 1972, unless otherwise noted.

U = Overall heat transfer coefficient

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1. CONCLUSIONS

The effectiveness of the Esso oil concentration process for concentrating secondary sludges, and mixed primary plus secondary sludges, has been demonstrated by both laboratory and pilot plant tests. Sludge concentrations of up to about 9% solids have been attained, including corrections for solubilized sludge solids and for the oil soluble component of the feed sludge (12% solids on uncorrected basis).

Cost estimates show a considerable advantage over the best comparable commercial technology (gravity thickening of primary, air flotation thickening of secondary, vacuum filtration, incineration). This advantage is from 13 to 31 $/ton dry solids for a 189 ton solids (primary plus secondary) per day plant. The cost advantage is larger for smaller plants.

Treatment costs for the Esso-Carver Greenfield process (including depreciation) are estimated at 21 - 39 $/ton dry solids for the 189 ton/day plant treating primary plus secondary solids. These cost are based on waste secondary sludge feed at 0.5% solids, with the higher numbers corresponding to a sludge very difficult to process. The economics include a debit for the cost of BOD recycle for any solubilized solids.

The Esso concentration step couples very well with the Carver Greenfield evaporative drying process; the oil used for the concentration provides the oil required in the evaporation to maintain fluidity and high heat transfer. The oil soluble component of the feed sludge serves as a partial replacement for the oil burned with the dry sludge solids in the incineration step.

An extensive laboratory study was first made in order to identify the factors controlling the concentration process and in order to optimize the process response. The degree of concentration obtained for a given sludge was primarily a function of the time and temperature of settling after the contacting of the oil plus the sludge; concentration increased with increasing time and temperature over the range covered, which was up to 90°C and up to 70 hours. Solids content after concentration was increased about 20% by adjusting the pH of the sludge feed from the initial near neutral pH to 3.0.

Little or no effect of oil type, oil/sludge ratio, and initial sludge solids content on final concentration was found. Use of surfactants, covering a wide range of HLB* values and chemical type, produced only a slight increase (if any) in solids concentration.

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* See Glossary, following Table of Contents

High mixing intensity as measured by shear rate and exemplified by a centrifugal pump, was found necessary to obtain $\geq$ 98% capture of the feed solids over the full range of feed solids contents tested. With relatively low shear mixing, such as produced by a turbine impeller, the solids captures were as low as 60%. A centrifugal pump is therefore a practical, low cost mixer for use in scale-up if needed.

Solubilization of some of the organic matter from the sludge solids has been observed to increase with the temperature and time of the oil-sludge thickening step. Some of the decomposed material also distills over into the drying stage of the evaporator. The economics in this report include a penalty for 25% BOD recycle, although reduction to 10% is believed possible with lower temperature settling.

Further possible net economic improvments to the proposed process totalling from 1 to $12/ton dry solids have been identified by use of a lower settling temperature.

2. RECOMMENDATIONS

Based on the experimental data available from a laboratory process study and the Carver Greenfield heat transfer tests, plus the results of a process and cost analysis, the route to a considerable reduction in the cost of the Esso Carver Greenfield process can be clearly predicted and the savings projected. In addition, several areas requiring further work to either explore potentially promising leads or better define the process response for different sludges have been identified. This will enable further pilot plant work to be done on the most favorable process system. Finally, evaluation of the Esso process component in a continuous pilot plant, with several different sludges is considered

To carry out the work needed to meet the objectives indicated above a program is recommended with the following specific objectives:

2.1 Confirm Projected Improvements in Present Process

- Reduce total TC loss in both concentration and evaporation steps, and increase overall heat transfer coefficient (U) value for evaporation, by operating the concentration step at a lower temperature.
- Reduce design area of oil-sludge settlers 50%, by confirming less conservative factor for scale-up to commercial sized plants.
- Reduce design area of sludge thickeners by obtaining firmer scale-up data than obtained from the static 1 liter batch settling tests.

2.2 Increase Concentration Factor and/or Increase Rate of Concentration

- Evaluate effectiveness of polyelectrolytes and new additives.
- Determine relationship between mixing intensity (shear rate) and time) with concentration factor and concentration rate.
- Explore potential of combined oil extraction-air flotation system.

2.3 Establish Firmer Basis for Range of Process Response for Different Sludges

Only 3 different sludges were tested and at least one is believed to be a very "difficult" material to concentrate. A considerably wider range of sludge samples is therefore required for a better measure of the variation in process response.

There is good reason to believe that 1) sludges from some plants will concentrate to higher solids contents than attained in this current program, 2) most sludges will concentrate to the higher levels attained in our study.

The operation of a continuous unit on site is the only practical way to be sure of the satisfactory scale-up of the mixing and settling steps. Successful operation of a continuous prototype unit at plant sites will hopefully also serve as a confirmation of a novel process concept to careful and conservative plant management.

3. INTRODUCTION

The handling and disposal of sludge from sewage plants has often been called the most troublesome aspect of the entire treatment process (1, 2). The volume of sludge requiring disposal is enormous, as can be inferred from the estimate that the average daily production of sludge is about 0.2 tons per 1000 people on a dry basis (9). Not only is sludge disposal a major problem now, but it is a growing one; within the next 15 years the volume of sludge will increase by an estimated 60-70% (10). Finally, disposal is a very costly operation, representing up to 50% of the total capital and operating costs of the treatment plant (2).

Because of the technical and economic importance of the problem, our country and other industrialized countries have become increasingly aware of the need for development of more efficient and lower cost processes to accomplish this disposal operation. With the rapidly growing awareness and concern for environmental protection, however, all new approaches must be geared to the objective of disposal without causing damage to the environment. As stated by the Chicago Metropolitan Sanitary District (1), the ultimate goal of any solution must include the following elements:

- Low cost
- Not produce air, water or land pollution.
- Make beneficial use of the sludge constituents.
- Solve the problem in perpetuity.

There is no lack of commercial processes for treatment and disposal of sludge with the most widely used listed below:

- Thickening - initial volume reduction
  - Gravity settling
  - Centrifugation
  - Air Flotation
- Stabilization - mass reduction
  - Anaerobic digestion
  - Aerobic digestion
  - Wet air oxidation
  - Heat treatment

- Dewatering - further volume reduction
  - Vacuum filtration
  - Pressure filtration
  - Centrifugation
- Chemical conditioning usually used for dewatering processes
- Ultimate Disposal
  - Dumping at sea
  - Soil conditioning and/or fertilization
  - Land fill
  - Incineration

Any of the treatment processes can be combined with any of the disposal methods. Each individual process and process combination has technical and/or economic problems and drawbacks. In the opinion of many, incineration is the only practical long term solution for sludge disposal for large and growing urban areas (11, 54). This view tends to gain credence in light of a) the increasing questioning of and restrictions on dumping at sea, b) the rapid decrease in available acreage for land fill, and c) the limited practical outlets for use of sludge as soil condition/fertilizer (1,2,9,10,11).

All treatment + disposal processes have one technical problem in common: the necessity of handling the very dilute sludges produced by the sewage plant. Primary sludges from the sedimentation tanks normally have a concentration of 2.5-5% with activated sludges 0.5-1.0%; these concentrations represent water/solids ratios of 200/1-20/1. Irrespective of the combination selected for treatment and disposal, there is a large economic incentive for dewatering of the sludges prior to processing, in order to reduce the volume that must be processed and/or to minimize the quantity of water to be removed during processing (2,4,9).

In considering sludge disposal a distinction must be made between primary + secondary sludges and digested sludges. Digested sludges, at least partially stabilized in regard to further decomposition, may be amenable to disposal techniques such as land fill and dumping at sea, which are not open to the raw or unstabilized primary + secondary sludges. Development of an improved process for disposal of the primary + secondary sludges is of prime concern for two reasons: the disposal problem is greater because of more limited disposal options, and an economically attractive process could eliminate the need for treatment by digestion.

Secondary sludges either from activated or trickling filter processes are the most difficult to concentrate prior to subsequent treatment steps. Current commercial concentration processes can be classified as "thickening" or "dewatering" according to the fluidity of the product (2);

Thickening	Dewatering
Gravity	Vacuum Filtration
Air Flotation	Centrifugation
Centrifugation	Pressure Filtration

These processes are normally used to concentrate separately either primary or secondary sludges, or mixed primary + secondary sludges. All of these processes have limitations in the solids concentrations attainable and/or the cost. Assuming that incineration will in the future be the preferred disposal method, there is an incentive for the development of a new, lower cost process for the total, combined dewatering + disposal process.

Esso proposed a new process to accomplish the stated objective; this process consists of combining a novel Esso sludge dewatering technique, based on an oil activated concentration step, with the commercial Carver Greenfield multiple effect evaporative sludge drying system. The Esso process component concentrates the sludge feed to a specified level; the Carver Greenfield process component completes the dewatering and produces a dry feed suitable for incineration in a conventional incinerator-boiler. The energy recovered from the burning of the sludge is reused in the evaporation step to provide maximum energy efficiency. The Carver Greenfield evaporation system is based upon use of a water insoluble oil as a fluidizing carrier for the sludge solids; this maintains high heat transfer rates even at very low water contents and prevents fouling of the heat exchange surface. The oil required for the Esso concentration provides the oil required for the Carver Greenfield evaporation step; the separate concentration and evaporation steps therefore dovetail extremely well into an overall integrated process. A patent application has been filed on the Esso process based on work done before the contract; the Carver-Greenfield process is patented.

Preliminary cost estimates for the proposed Esso Carver Greenfield process were very attractive, with total operating costs/ton sludge solids of ≤$40 for plant sizes of ≥20 tons/day. The present contract was undertaken in order to a) make a detailed variable study in order to optimize the dewatering process, b) to develop the process costs for a range of operating conditions and options on the basis of the experimental data obtained in the study.

The contract was carried out as a four part program:

Phase 1: Laboratory Process Development and Optimization.

Phase 2: Pilot Plant Scale-up

Phase 3: Heat Transfer Studies at Carver Greenfield

Phase 4: Process Trade Off and Cost Analysis

In the first two phases, concerned with the study of the controlling process parameters, the process responses evaluated were the solids concentrations achieved, the degree of capture of the feed solids and the quality of the water phase (defined as the "raffinate") separated during concentration. These latter two factors were important in determining the recycle load generated by the process. Recycle load is an important consideration, since one of the projected advantages for the Esso Carver Greenfield process over several current processes (heat treatment, wet oxidation, centrifugation) was low recycle.

The initial theoretical basis for the Esso process was the selective "wetting" of the lipophilic sites on the sludge solids by an oil with properly matched properties, followed by "flotation" of the oil droplets with attached sludge solids to form a concentrate phase. One component of the laboratory process study was to attempt to confirm this hypothesis and to follow technical implications derived from the initial model.

4. PHASE 1: LABORATORY PROCESS DEVELOPMENT AND OPTIMIZATION STUDY

The objectives of the laboratory program were to a) evaluate the effects of the variables considered to be potentially important for the oil concentration process, b) establish the operating conditions required to "optimize" the process in terms of the different performance criteria set up and, c) develop a range of process alternatives (derived from a and b) required for the final prototype commercial designs and the overall process cost analysis.

4.1 Experimental Program and Procedure

4.1.1 Variables Tested

Over the course of the laboratory program an extensive list of variables was evaluated; the variables and the range tested for each are summarized in Table 1. The variations in response of the different sludge batches of the same type from the same plant, and the limited "shelf life" of the batches can, in a practical sense, be considered implicit variables; problems associated with these two factors imposed severe restrictions at times on the design of the test program.

In the discussions below on the effects of specific variables, the order presented does not necessarily reflect the chronological order in which the work was actually done, nor the relative importance of each variable. Because of the batch/batch variability, individual experimental results are identified by batch for convenience. In many cases a specific experiment will appear in more than one summary (or tabulation) for convenience in making comparisons where the data are pertinent to more than one variable parameter.

4.1.2 Experimental Procedure

The sludge sampling, transport and storage procedure, as well as the description of the test procedure, are described in detail in Appendix A-1; a condensation of this procedure follows:

- Sludge was kept stored in refigerator at $40^{\circ}F$ except when removed to obtain material for the day's tests.
- A measured quantity of sludge, adjusted to the required temperature and pH was charged to the mixing unit with a measured quantity of oil preheated to the required temperature.
- After mixing for the specified time, where batch mixing with a turbine agitator or Waring Blender was used, the combined oil-sludge was transferred for settling to calibrated 250 or 500 cc straight sided glass settlers

TABLE 1

EXPERIMENTAL VARIABLES TESTED IN LABORATORY PROGRAM

Variable	Range Tested
● Sludge Type	
- activated	2 different plants, total of 13 different batches.
- trickling filter	1 plant, total of 2 different batches.
- primary + activated	2 plants, total of 3 different batches.
- digested	1 plant, 1 batch
● Sludge Properties	
- initial suspended solids content	0.5 - 4.6%
- initial pH	3.0 to 6.5-7 (as received)
- initial temperature	8-50°C (46-120°F)
● Mixing Intensity (Shear Rate) and Time	hand mixing, turbine impeller, centrifugal pump, Waring Blender
● Oil Type	6 oils
● Oil/Sludge Ratio for Extraction	0.1 - 0.8 (volume basis)
● Concentration (Settling) Temperature After Mixing	25 - 90°C (75 - 194°F)
● Use of Surfactants	HLB of 1-10, 100-100,000 ppm based on solids, several chemicals types.
● Effect of Impurities in Recycle Oil	Recycle oil from commercial plant, simulated recycle subjected to "degradation" for varying times.

and stored in constant temperature ovens. For pump mixing, the oil + sludge was premixed in a stirred vessel and fed thru the pump at a controlled rate into the settlers.

- **The water raffinate-oil sludge interface level was periodically measured and the degree of concentration calculated (see below for procedure used). Alternatively, the water raffinate phase was withdrawn and weighed.**
- Samples for Total Carbon (TC) analysis were submitted "as is" from raffinates without visible settled solids (from pump and Waring Blender runs); samples containing settled solids, from feed sludge or where turbine mixing was used, were first centrifuged.

4.1.3 Parameters Measured in Laboratory Tests

Very early in the program it became clear that three different factors had to be measured to adequately describe the test results:

- Increase in sludge concentration vs. settling time: calculated from the initial weight of sludge, and the weight of water raffinate phase separated

$$\text{Concentration Factor} = \frac{\text{\% Solids in Concentrate (Water Phase Basis)}}{\text{\% Solids in Feed Sludge}} = \frac{\text{Initial Sludge Volume}}{\text{Initial Sludge Volume} - \text{Raffinate Volume}}$$

- Solids "capture" in oil-sludge phase: determined by filtering the raffinate phase and weighing the solids.
- **Decomposition/solubilization of sludge solids: calculated from total organic carbon and/or total carbon analysis of initial sludge solids, and of the water raffinate phase.**

4.1.4 Basis for Selection of Sludge Sources

The contract specifies that the test sludges include two different activated sludges, one trickling filter secondary sludge, and one mixed primary and secondary sludge. The most important considerations were that the plants should be representative of the important secondary treatment plants in the country and adequately convenient to the laboratory. Based on a review of the sewage plants in the New Jersey area with secondary treatment, Bergen County, N.J. and Wards Island, NYC were selected as activated sludge sources and Trenton, N.J. as trickling filter source. General information on these sources is summarized below.

4.2 Analysis of Sludges and Plant Streams

4.2.1 Waste Sludges from Sewage Plants

Chemical analyses and respiration rates obtained on the early samples taken from the plants are presented in Appendix A-2 and analysis of the dry sludge solids filtered from the waste sludges in Table 2 (page 13).

Results obtained for the dry sludge solids and the oil fraction are in line with data reported in the literature (7, 29).

Data on the hexane extractable component of the total sludge feed (defined in the rest of this report as "oil") are shown in Table 3.

TABLE 3

HEXANE-SOLUBLE CONTENT OF SLUDGES

Sludge Batch (1)	Type	Oil Solubles Wt. %
Bergen County	activated	9.5
Bergen County	activated	16.7
Wards Island	activated	8.2
Bergen County	primary + activated	6.2
	Average	10.2

---

(1) Sludges used for Phase 3 program, in order processed.

This oil component, which averaged 10% for the 4 batches tested, is an important factor in calculating the overall heat balance for the system and is required to calculate the effective solids concentration going into the Carver-Greenfield evaporation step; since the oil component of the feed will dissolve in the process oil used for the concentration step, the initial solids content must be reduced in calculating a) the true solids concentration achieved and b) the solids load for incineration.

4.2.2 Plant Treatment Data

Analytical results obtained from the plant laboratories for the various plant streams are assembled in Appendix A-3, 4, 5.

TABLE 2

ANALYSIS OF SLUDGE SOLIDS DRIED AT 102°C (1)

Batch	Type Sludge	% Volatile at 1000°F	Elemental Analysis C	H	N	P_2O_5
Bergen County						
A	Activated	63.0	29.7	4.7	5.3	
B		63.1				
C		64.6	32.8	4.6	6.1	
D		68.8	33.0	5.2	6.5	
E		66.3	34.2	4.9	5.9	
F		68.6	34.8	5.3	6.0	3.1
G		66.0	34.1	5.1	6.0	
J			37.5	5.6	6.3	
D	Primary + Activated	65.1	34.0	4.8	3.4	
E		63.3	31.0	4.6	4.3	
I		71.4	37.9	5.5	4.9	1.6
D	Digested	55.3	31.9	4.7	4.7	
Wards Island						
A	Activated	70.1	36.7	5.2	6.6	
B		69.6				
D		68.7	38.9	5.8	5.2	
Trenton						
A	Trickling Filter	39.3 (2)	20.9	3.3	2.7	
B		50.1	27.6	4.1	2.5	

Bergen County I - Qualitative Analysis of Ash (after volatiles removed)

Wt. %			
≥ 10	Al	.1-1	Ba, Sr, Mn, Mg, Zn, Ti
1-10	Ca, Cr, Fe, P	<.1	B, Pb, Sn, Mo, V, Cu, Ag

---

(1) All analyses by Analytical and Information Division of Esso Research

(2) Low values due to high rust content in sample.

TABLE 4

DESCRIPTION OF PLANTS PROVIDING SLUDGE SAMPLES

Plant	Type	Design Capacity MGD	Type Feed	Population Served (Approx.)
Wards Island, NY	Activated Sludge	220	Sanitary + industrial	750,000
Bergen County, NJ	"	50	"	250,000
Trenton, NJ	trickling filter	20	"	150,000

4.3 Why Does the Esso Oil Concentration Process Work?

A very brief review of the background used to initially develop the concept for the process, as well as the more recent modifications required by the actual test data, should provide a useful background for the experimental program.

The chemical and physical properties of activated sludge have been extensively described in the literature, (3, 12-16). Considerably simplified, the sludge solids can be considered as highly hydrated bacteria, bacterial fragments, and slimes produced by the bacteria, in a fine particle size, floc-type structure. These solids possess a very high surface area, with a large fraction of hydrophilic (hydrogen bonding) surface sites. The particle surfaces tend to acquire a negative electric charge. Ionized solubles and water will be attracted to and held to the surface of the particles by both of these sludge surface characteristics. The net result is a system of sludge solids with a strong affinity for water, a low effective specific gravity, and with a tendency to remain dispersed due to electrostatic repulsion (zeta potential). Chemically, the solids consist of a complex mixture of polysaccharides, proteins, amino acids, sugars, carbohydrates and high molecular weight polymers.

In our initial concept for the concentration of sludge solids by use of oil, the lipophilic sites on the solids were assumed to be wet by the oil, so that the solids transferred to the interface with the oil droplets; according to this concept, "free" water is expelled as a "raffinate phase", with mainly "bound" or hydration water remaining with the solids in the concentrate phase. The concentration process could be considered as analogous to solvent extraction, with oil the "solvent" and solids the material being "extracted" from the aqueous phase. On a more theoretical level, two unrelated steps were assumed to be involved:

1. Since the solids are collected by the oil, it is evident that the hydrocarbon has a contact angle less than that for complete non-wetting. Due to the omnipresent, non-selective London forces that exist among all molecules regardless of type, this is not 180° but only about 110°. Evidently, the lysed bacteria retain some waxy coating or other lipophilic spots which lower this to below 90°, so that the cosine of this angle has a positive value. This causes the finely divided solid to serve as an emulsifier and separate the oil into drops whose diameter is that of the solid particle divided by the cosine of the contact angle (50).

2. Since this cosine is not very large, the drops will be large and the emulsion of poor stability. However, solids-stabilized emulsions do not fail by coalescing, as the solids coating prevents oil/oil contact; the result is failure by creaming (17). This brings all the oil, all the solids and a minor amount of water into a layer which can be skimmed.

The above concept was put in considerable doubt after finding that the oil concentrate phase was actually an oil in water (o/w) emulsion. This fact was established experimentally by electrical measurements, showing that the specific resistance of the oil-sludge concentrate phase was equal to that of the sludge alone.

Based on literature correlations for o/w emulsion properties, the Hydrophylic-Lipophylic Balance (HLB) of the sludge system was estimated to be 12-14 and the Cohesive Energy Ratio (CER) 0.3-0.6 (18). Applying the calculation method of group contributions (18) on the assumption of a predominately cellulose structure gave values of 14 and 0.43 respectively, in excellent agreement with the estimate.

In the model of the system, the sludge solids were assumed to be attached to the rising oil droplets by simple contact adhesion; good mixing was assumed necessary both to insure contacting and capture of all the sludge solids by the oil, and to provide the needed energy for adherance. As an extension of this hypothesis, the lipophilic character of the solids should be enhanced by proper choice of the specific oil used and by adding an appropriate surfactant. As will be discussed in later sections, however, experiments along the lines of this hypothesis failed to produce the expected results, casting doubt on the basic concept.

Microscopic examination of several oil sludge concentrates showed no evidence of any solids adhering to the surface or even trapped within oil droplets; the smaller oil droplets or larger drops and globules formed by coalescence were suspended in the aqueous sludge, but untouched by any solids. Unless we assume that the solids separated from the oil droplets almost immediately after mixing, the physical adherance theory does not appear to be valid.

The current hypothesis is that the rising oil droplets are trapped by the interconnected, web-like floc structure of sludge solids and actually float them "en-masse". An analogy would be a blanket being floated by a number of balloons. An alternative possibility is that the oil droplets become trapped within floc masses. In either case, the individual solids particles do not physically adhere to the oil droplets.

4.4 Results of Process Study

4.4.1 Short Term Storage of Fresh Sludge Not Detrimental

Storage characteristics are of little importance in any commercial process, but initially were of considerable concern since the laboratory is removed from the plant sites. A further refinement of this problem was the uncertainty about the need to refrigerate samples during transfer from plant to laboratory or pilot plant; total transfer time was in the range of 2-4 hours. Long shelf life (up to 1 week) was naturally hoped for in order to be able to minimize the frequency of procuring new batches.

Tests carried out at the start of the program (see Table 5) did not show any apparent effect in terms of response to the oil concentration process for 3-7 day storage at 40°F or 1-2 day storage at ambient temperature (80°F). Samples stored for 7 days showed a definite change in response, with the solids capture for mild agitation sharply lower. These results have been confirmed by more recent data with high shear agitation (Table 6). On the basis of the above, storage life at 40°F for laboratory work was set at 6 days maximum, with 5 days preferred, and 2 days at ambient for subsequent pilot plant work.

4.4.2 Solids Concentration Increases With Increasing Settling Time

Immediately after the mixing of the sludge and oil, a water raffinate phase starts to separate out of the oil-sludge mixture. The phase interface is well-defined and stable, provided the proper processing requirements of oil/sludge ratio and mixing intensity are met (these points are discussed in detail in the section below). After the initial very rapid breakout of the two phases, the rate of further separation of water decreases rapidly with time. The extent of solids concentration achieved as a function of the volume of water separated (raffinate) has been previously defined as:

$$\text{Concentration Factor} := \frac{\%\ \text{solids in concentrate (ex oil)}}{\%\ \text{solids in feed sludge}} = \frac{\text{Initial Sludge Volume}}{\text{Initial Sludge Volume} - \text{Raffinate Volume}}$$

TABLE 5

EFFECT OF SLUDGE STORAGE TEMPERATURE AND TIME ON CONCENTRATION RESPONSE

Sludge Batch	Storage Temp °F	Date of Test	Concentration Factor (2) For Settling Times of 1 Hr	5 Hr	16/20 Hr
Bergen County "A"(3)	40	8/10/71(1)	1.7		2.4
		8/11/71	1.6		2.7
		8/17/71	1.7		3.0
"A"(3)	40	8/12/71	2.5		4.6
		8/13/71	2.9		4.6
		8/16/71	2.9		5.5
"B"(3)	80	8/18/71(1)	3.3		7.1
		8/19/71	3.5		7.1
		8/20/71	3.3		6.6
"B"(3)	40	8/20/71	1.4		2.8
		8/23/71	1.5		3.1
"J"(4)	40	2/24/72(1)		7.7	10.6
		2/29/72		7.9	10.0
"K"(4)	40	3/6/72(1)		7.5	10.9
		3/9/72		7.5	11.4

---

(1) Date sampled from plant.

(2) Values shown are averages of 2 tests on same day; test conditions were constant for comparisons within a sludge batch; test conditions were not the same for different batches.

(3) Low shear agitation.

(4) High shear agitation.

As concentration proceeds, the importance of each fixed increment of raffinate volume becomes greater, as shown below:

TABLE 6

CONCENTRATION FACTOR VS RAFFINATE VOLUME

Raffinate Volume % of Initial Sludge	Concentration Factor
20	1.25
40	1.67
60	2.50
80	5.00
90	10.00

Typical curves for concentration factor vs. settling time are shown in Figure 1; all runs were made under the same set of conditions (mixing, settling temperature, oil-type, etc). The linear relationship, using the semi-log type correlation, held up to about 20 hours settling for almost all test runs made, and in some runs even up to about 70 hours. The increase in concentration factor per unit settling time decreases rapidly for all runs. Detailed test data for different sludges, settling temperatures, feed solids contents are tabulated in Appendices A-6 and A-7.

The data presented in the curves illustrate several additional points:

- The slope of the settling curve (rate of increase in concentration factor per unit time) increases with decreasing initial sludge solids content.
- For the same initial solids content, activated sludge and mixed primary + activated show the same concentration characteristics.
- Test reproducibility is good, with the difference between duplicates $\leq$ 10%; this can be seen from the curves, where the different symbols represent duplicate batches.

4.4.3 Concentration Factor Increases With Increasing Settling Temperature

As would be expected in any separations process, temperature has a very considerable effect. For all sludges, concentration factor increased with increasing settling temperature over the range tested, which was 25°C - 95°C (see Appendix A-8, for complete summary); typical data are shown in Figure 2.

FIGURE I
CONCENTRATION FACTOR VS. SETTLING TIME
80°C SETTLING HIGH SHEAR MIXING

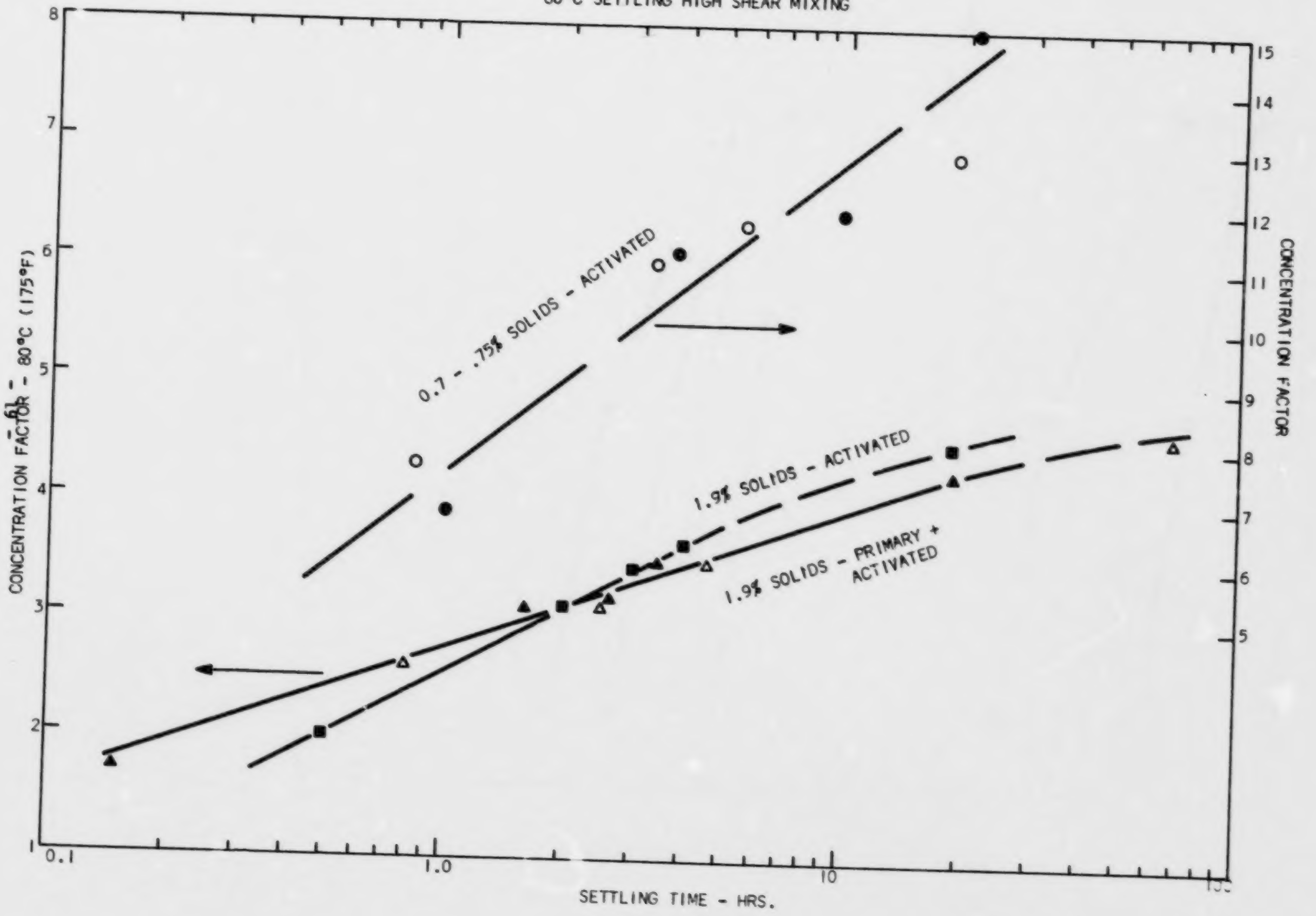
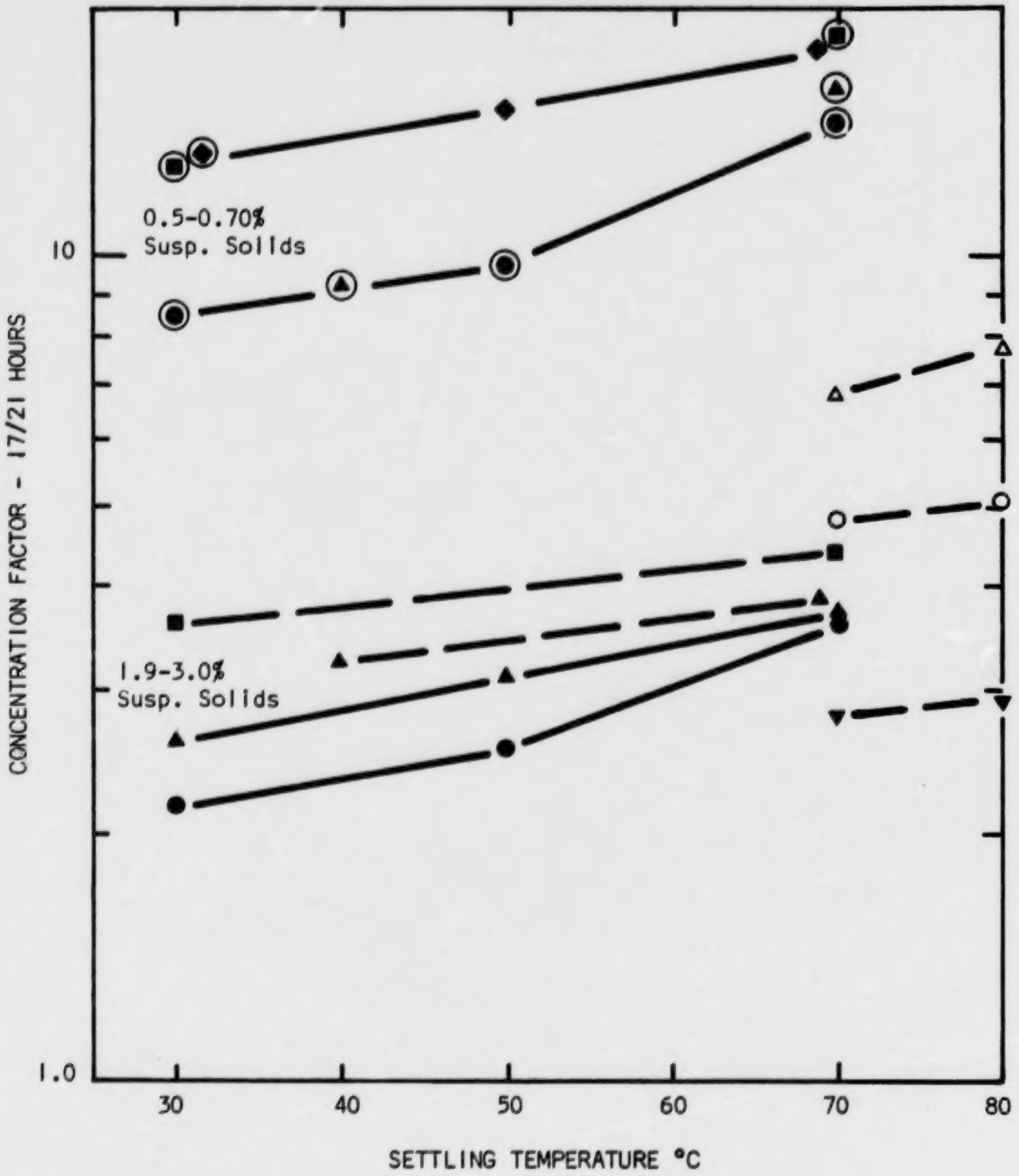


FIGURE 2

CONCENTRATION FACTOR VS. SETTLING TEMPERATURE (HIGH SHEAR MIXING)



The effect of temperature is inversely related to initial sludge solids content. For initial solids contents of 0.5-0.7%, concentration factor increases by 40-90% over the range of 45°C-80°C, vs. 10-40% increase for initial solids content of 1.7-3% for the same temperature span.

% Solids in Feed Sludge	Concentration Factor Ratio 80°C/40-50°C (1)
0.5-0.7	1.54
1.7-3.0	1.19

(1) Difference significant at 97% confidence level.

Limited data for the range 80-95°C indicate relatively little further increase in concentration factor; the average increase was about 6% for initial feed solids contents of 1-2%.

The concentration process can be considered as a combination of "bulk flotation" of the sludge floc by the oil, sedimentation of the oil droplets, and coalescence of the settled oil droplets. Increasing temperature should therefore increase the rate of "flotation" by increasing the density difference between the oil and water phases, and the rate of sedimentation by reducing the viscosity of the continuous water phase. As shown below, the temperature effect on both density difference and water viscosity is substantial.

Temperature, °C	Water Viscosity Centipoises	Specific Gravities		Δ Sp. Gravity
		Water	#4 Heating Oil	
25	0.894	.997	.884	.113
40	0.656	.992	.870	
80	0.357	.972	.824	
95	0.299	.962	.819	.143

Settling temperature is one of the important trade-off factors considered in the final process and cost analysis phase of this project. While increasing temperature does increase solids concentration, balancing factors are the added cost for the large quantity of extra heat required and the adverse effect on sludge solubilization/decomposition.

4.4.4 Centrifugal Pump is Satisfactory and Practical High Shear Mixer

Based upon early test (Appendix A-9) data, only a very high shear mixer, the Waring Blender, was suitable for extraction of low (<0.7%) suspended solids content sludges. Even for settled sludges, the solids capture for the Waring Blender was better than for the mixing type turbine. Since a Waring Blender cannot be scaled up to commercial size, a variety of other practical mixer types, capable of generating high shear and of scale up to the large commercial size required, were considered. An additional factor was to find a mixer which could also be used in the pilot plant operation.

A standard centrifugal pump was found to be very effective; solids capture and concentrations equal to the Waring Blender and superior to the turbine were achieved, as summarized in Table 7 below and detailed in Appendix A-10 and A-11.

TABLE 7

CENTRIFUGAL PUMP IS SATISFACTORY MIXER

Type Sludge	% Solids in Feed	Type Mixing	Concentration Factor 1 Hour	Concentration Factor 20 Hours	% Solids Capture
Activated	1.0	turbine			~80
		centrifugal pump	3.2	8.3	98
	1.8	turbine	1.6	3.2	95
		centrifugal pump	2.8	4.8	98
Primary and Activated	1.9	turbine	2.1	3.8	95
		Waring Blender	2.9	4.4	98
		centrifugal pump	2.8	4.5	98

The specific pump used in the laboratory program was a 1/20 HP, 6000 RPM, single stage, open impeller, Eastern Industries Company pump.

A centrifugal pump can be used on a commercial scale, either conventionally or with reverse flow feed for greater mixing efficiency if required.

From a limited amount of testing, excessive mixing in the pump appears detrimental to achieving maximum concentration. Increased mixing was produced by use of a second mixing pass through the pump. The residence time (mixing time) in the pump was ~0.3 seconds per pass. This result is directionally consistent with the data for the Waring Blender, where concentration factor was related to mixing time.

% Solids in Feed	Number of Passes through Pump	Number of Test Runs	% Solids in Concentrate After Settling 1 Hour	22 Hours
1.0	1	2	3.5	9.0
	2	2	3.3	5.6
1.	1	1	3.1	5.4
	2	1	3.4	4.5

The relationship between mixing intensity and solids capture can be put on a more quantitative basis by defining mixing in terms of shear rate:

$$\text{Shear rate, sec}^{-1} = \frac{\text{Impeller tip speed, cm/sec}}{\text{Clearance between impeller tip and mixing chamber wall-cm}}$$

For all sludges tested solids capture increases with increase in shear rate; the sensitivity of the solids capture to shear rate decreased with increasing sludge solids content, as is summarized below:

TABLE 8

MIXING SHEAR RATE CONTROLS SOLIDS CAPTURE

Type Sludge	Type Agitation	% Solids in Feed	Shear Rate sec^{-1}	Solids Capture - %
Activated	Turbine	0.5	75	~60
	Turbine	0.8	130	80-90
	Turbine	1.5	130	95
	Turbine	2.3	130	98
Activated	Centrifugal pump	≥ 0.5	73,000	98+
	Waring Blender	≥ 0.5	210,000	98+

The mixing intensity for the turbine agitator - baffled vessel system would be considered vigorous for a batch mixing system, but is low compared to either the pump or blender. Some intermediate shear rate between the batch turbine and centrifugal pump probably will be adequate to insure high solids capture, but remains to be defined.

4.4.5 Satisfactory Solids Concentration Achieved for Different Type Sludges

Activated, mixed primary + activated, and digested sludges from the Bergen County plant were all successfully concentrated with high solids capture using high shear mixing (see Appendix A-11). Results for 20 hour settling at 80°C with a Waring Blender are summarized below for comparison, with the final solids content adjusted for average % oil solubles in the feed sludge and for average TC losses in the raffinates:

TABLE 9

OIL CONCENTRATION PROCESS WORKS FOR DIFFERENT TYPE SLUDGES

Type Sludge	% Solids in Feed	% Solids in Concentrate
Activated	2.3	8.5
Primary + Activated	2.7	9.6
Digested	2.8	6.9

While the digested sludge was not included in the original program, a very limited amount of work was considered desirable to demonstrate the suitability of the Esso process for the full range of sludge types produced in sewage plants.

4.4.6 Final Solids Concentration Not Affected by Initial Feed Solids Content

As noted above, for a given sludge type and source the concentration factor (a measure of the rate of separation of water phase) is inversely related to the initial feed solids content: concentration factor decreases with increasing feed solids content; this is shown in Figure 3, for all test data at 80°C. The final solids concentrations achieved, however, appear to be independent of initial feed solids content, using the data from Figure 3 and Appendix A-12; the ranges of final solids contents for different feed solids contents are summarized in Table 10:

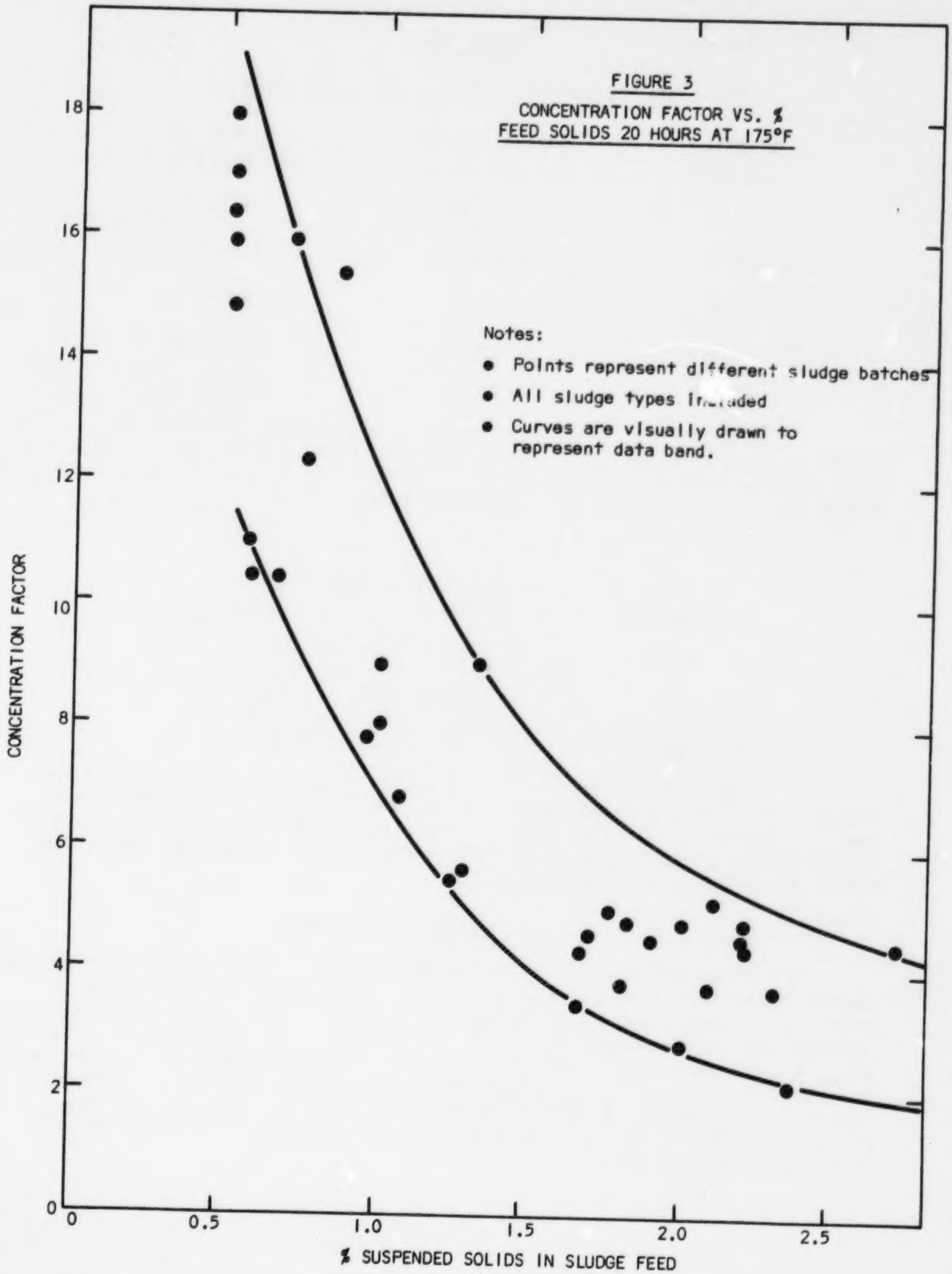


FIGURE 3

CONCENTRATION FACTOR VS. % FEED SOLIDS 20 HOURS AT 175°F

Notes:

- Points represent different sludge batches
- All sludge types included
- Curves are visually drawn to represent data band.

TABLE 10

FINAL SOLIDS CONCENTRATION VS % SOLIDS IN FEED

% Solids in Feed	Concentration Factor	% Solids in Concentrate
0.5	12-20	6-10
1.0	6.5-11.5	6.5-11.5
1.5	4-8	6-12
2.0	2.8-5.9	5.6-11.6
2.5	2.2-4.7	5.5-11.7

This conclusion is based on the generalized curve and is supported by the results for individual sludge batches tested over a range of feed solids contents; data for one batch, Bergen County activated sludge LF-"J", settled at 80°C, are summarized below:

% Solids in Feed	% Solids in Concentrate	
	5 Hrs	20 Hrs
0.70	7.8	10.0
0.75	8.0	9.6
1.0	7.7	10.5
1.5	7.8	10.3
3.0	7.7	10.3

This lack of effect of feed solids content on final solids concentration is an important factor in the overall process analysis and cost estimation. The trade off factors will be limited to the Esso oil concentration process components, since the solids content to the Carver Greenfield process will be constant.

4.4.7 Oil/Sludge Ratio for Concentration Step

Since the oil concentration process is coupled directly with the Carver Greenfield process, the oil/solids requirement for the Carver Greenfield process can be considered as a potentially limiting factor; the minimum oil/solids weight ratio for the Carver Greenfield process is about 10/1, with a preferred range of 10/1-15/1. Use of a lower oil/solids ratio for the Esso concentration step presents no problem (oil can be added before the evaporation), but a higher ratio would be undesirable; the Carver Greenfield requirements therefore set the preferred range for the Esso concentration step.

As shown in Appendix A-13, the oil/sludge (O/S) ratio required to provide the preferred oil/solids ratio is a function of the initial feed solids content of the sludge; this varies from a ratio of about 0.06 for 0.5% initial feed solids content to about 0.25 for 2.0% feed solids content, on a volume basis.

For evaluating the effect of oil/sludge ratio on the concentration step itself, a range of 0.1-0.6 was used for most of the testing. Considering only the effect of oil/sludge ratio on concentration factor, the optimum response was obtained at O/S values of 0.1-0.2 for nine out of the ten batches evaluated (Appendix A-14). On average, the concentration factor descreased with increasing oil/sludge ratio:

Oil/Sludge Ratio	Concentration Factor	
	1 Hour	20 Hour
.1	2.8	5.0
.2	2.5	4.6
.4	2.2	4.4

With an oil/sludge ratio of 0.1, however, the interface between the oil + sludge concentrate phase and the water raffinate phase was less sharp and less stable than at higher ratios. Slight movement of the settler caused sludge solids to disengage from the concentrate phase and settle in the raffinate. From practical considerations of interface stability and requirements for the Carver Greenfield process, thus, an oil/sludge ratio of 0.2 appears to represent the best choice; this value was therefore used in the design basis for a commercial plant in Phase 4.

The majority of the tests in the laboratory program were made within an oil/sludge ratio of 0.2, and almost all with the O/S ratio 0.2-0.4. The effect of O/S ratio on any of the results for the other variables was therefore very small, if any.

4.4.8 #4 Heating Oil Preferred Oil for Sludge Concentration Step

A wide variety of candidate hydrocarbon oils are available for the extraction process. The oils have different physical properties (density, boiling point, viscosity) and chemical composition (paraffin/aromatic content). The oils selected for the screening study and their properties plus approximate cost, are summarized in Appendix A-15.

Essentially all of the comparisons of oils involved #4 and #2 Heating Oils, #1 Varsol and LOPS; these comparisons, which were carried out for different sludges and % feed solids contents, as well as different mixing systems, are tabulated in Appendix A-16. While differences of up to about 15% where found between oils for individual runs, there were no apparent consistent differences. On average, #4 Heating Oil was as good as any of the other oils in terms of concentration factors obtained. Of the four oils most completely evaluated, #4 Heating Oil, #1 Varsol and LOPS appeared about equal in performance; concentration factors with #2 Heating oil were slightly inferior.

The performance of #4 Heating Oil was at least as good as the other candidate oils for the sludge concentration and has the lowest cost; this oil was therefore used for the bulk of the process studies.

TABLE 11

COMPARISON OF OILS FOR CONCENTRATION

Oils Compared	No. Tests in Comparison	Ave. Concentration Factor (20 Hour Settling)
#4 Heating Oil #1 Varsol	7	3.8 3.7
#4 Heating Oil LOPS	3	3.1 3.0
LOPS #2 Heating Oil	2	4.0
#4 Heating Oil #2 Heating Oil	2	3.5 3.1

The initial "model" for the oil concentration process involved the following sequence:

a) The lipophilic sites on the sludge solids were supposed to be "wet" by the oil.

b) Followed by "flotation" of oil droplets with sludge solids adhering to the droplet surface.

c) The oil droplets with adhering solids then formed a concentrate phase. Proper selection of the oil, on the basis of matching the oil properties to the sludge solids surface using three dimensional solubility parameters was considered essential for maximizing solids concentration. The wide range of oil types tested in this study was based on the logic of the initial model.

Subsequent microscopic examination of the oil sludge concentrate showed that the model above was not valid, that the sludge solids did not adhere to the surface of the oil droplets due to any "wetting" phenomena. The chemical composition of the oil and its derived interfacial properties should have no effect on the concentration and this is what the tests actually showed.

The revised "flotation" model, however, implies that at least the rate of concentration, if not the final concentration achieved, should be a function of the oil density; rise velocity of a given size oil droplet is proportional to the density difference between the oil and the bulk phase fluid (water). The lack of any apparent oil effect even on the rate of concentration can be due to two factors:

- Sludge solids show a dominant effect so that the "flotation" rate is controlled by the rate of "escape" of the free water through the interlocking floc structure of the sludge solids.
- Oil droplet size interacts with density, so that the effect of lower oil density is offset by the formation of smaller droplet size.

4.4.9 Surfactants Have Little Effect on Concentration Factor

The initial concept of the oil concentration process assumed that oil adhered to the lipopohilic sites on the sludge solids, to effect transfer to the concentrate phase. Use of surfactants was considered an attractive possibility for increasing solids concentration by increasing the "wettability" of the sludge solids by the oil. A wide range of surfactants, in regard to HLB and chemical type, as summarized in Table 12, were therefore screened for effectiveness (19). The surfactants were tested at two stages of the concentration process; during the initial concentration step, (when the sludge was mixed with the oil) and after the standard concentration had been completed. The surfactants tested were all oil soluble and added to the system by dissolving in the oil before mixing with the sludge or oil-sludge concentrate.

In the initial exploratory screening test, emulsification of the oil with water was found with surfactants having HLB $\geq$ 7.8 thus reducing the actual concentration obtained; an increase in solids concentration was found with a surfactant of the same chemical type, but with an HLB = 3.6. Based on this lead, a more intensive evaluation was carried out using surfactants with HLB values $\leq$ 3.6 in order to minimize the undesirable emulsification; these surfactants are strongly liphophilic in character. Several duplicate tests were carried out using different batches of fresh sludge and at both 25°C and 80°C; improvements in concentration factors for any given test were relatively small ($\leq$ 10%) and not consistent from test to test (see Appendix A-17 for representative data).

TABLE 12

SURFACTANTS TESTED WITH SLUDGE

Surfactant (1)	Type	HLB Value	Chemical Description
Triton X-15	Nonionic	3.6	Octyl phenoxy polyethoxy ethanol.
Triton X-35	Nonionic	7.8	Octyl phenoxy polyethoxy ethanol.
Triton X-35	Nonionic	12.4	Octyl phenoxy polyethoxy ethanol.
Atmos 300	Nonionic	2.8	Mono and diglyceride of fatty acids.
Span 85	Nonionic	1.8	Sorbitan Trioleate
Oleic Acid	Anionic	1.0	Representative of the acids in commercial Tall Oil Fatty Acid
Armoflo 49 (Armeen T)	Cationic		Primary aliphatic amine
Paranox 24	Anionic		Calcium sulfonate; MW of ~900

SURFACTANTS TESTED WITH OIL-SLUDGE CONCENTRATE

Surfactant(1)	Type	HLB Value	Chemical Description
Tween 81	non ionic	10.0	Polyoxyethylene sorbitan monooleate
Span 60	non ionic	4.7	Sorbitan monostearate
Span 40	non ionic	6.7	Sorbitan monopalmitate
Span 20	non ionic	8.6	Sorbitan monolaurate
ECA 4360(2)	cationic		Detergent cleaner type
Paranox 30(2)	anionic		Barium sulfonate
F-0525(2)	non ionic		Demulsifier, amine type

(1) Trade names for commercial products.

(2) Proprietary and/or experimental products.

Comparable results were obtained when evaluating the surfactants on the oil sludge concentrate, rather than the fresh sludge as discussed above; one additive, an experimental Enjay Chemical Co. demulsifier, F-025, produced a 10% additional increase in concentration, with no effect found for any of the others evaluated.

Considering the combined data for the different tests, the improvement in concentration that could be expected with any of the surfactants tested is no more than 10% and probably less. The first test series (fresh sludge) was conducted at a very high surfactant dosage. In the second series, the oil sludge concentrate dosage was varied from 50-10,000 ppm based on sludge solids; for the one additive with any beneficial effect, a dosage of ~1000 ppm was required for maximum effect. At this dosage the cost/ton sludge was estimated at ~$7 for the additive, assuming a "once-thru" basis (no reuse). Further work would be required to a) firmly establish the reuse factor for oil soluble additives and b) to evaluate other additives in the same structural type series which may be more effective. Unless considerably greater improvements in concentration factor than those found to-date were consistently obtained, or a large reuse factor confirmed, the economic incentive for surfactant (additive) usage appears to be small.

4.4.10 Lowering Sludge pH Increases Solids Concentration

The sludge solids carry a negative surface charge which produces a repulsive force between particles; this repulsive force is believed to contribute to the poor settling and compaction characteristics of secondary sludge. The surface charge can be neutralized by lowering the pH from the initial 6.5-7.0 to the isoelectric point, which is reported in the literature as occuring at pH 2-3 (16). Adjustment of sludge pH was therefore evaluated as a means of increasing solids concentration. Another reason for acidification is to "shrink" the proteinaceous solids by reducing the degree of hydration.

pH adjustment was tested at two levels, pH 4.0 initially and than at pH 3.0; test data are tabulated in Appendix A-18 for pH 4 and Appendix A-19 for pH 3, and summarized below in Table 13:

TABLE 13

ACIDIFYING SLUDGE INCREASES CONCENTRATION FACTOR

Sludge pH	Relative Concentration Factors (80°C)	
	1 Hour Settling	20 Hour Settling
Unadjusted (6.5-7)	1.00	1.00
4.0	1.20	1.10
3.0	1.33	1.15

Solids concentration, as measured by average concentraction factor, increased with decreasing sludge pH over the range tested. The effect of pH appeared to be greatest at the beginning of the settling step, but was still substantial after 21 hours at the termination of the test. Based on the three batches tested, solids content can be increased by 15% after 20 hours settling at 80°C and pH 3, compared to the unadjusted sludge.

As discussed in the sections on Total Carbon losses in the raffinate, pH adjustment has the added benefit of reducing TC losses. From the practical standpoint of commercial operation, these benefits must be balanced against the costs of pH adjustment: chemicals, added storage and mixing equipment, corrosion resistant settlers. This process option has been included in the Phase 4 Process and Cost Analysis.

4.4.11 Factors Controlling Loss of Feed Solids During Concentration Process

The effect of the oil concentration process on the quality of the water raffinate, in terms of dissolved solids from the feed sludge, was monitored by analysis for Total Organic Carbon (TOC) and/or Total Carbon (TC). The loss of feed solids in the raffinate is important for establishing a) if the raffinate streams require recycle, b) if so, the magnitude of the recycle load relative to plant capacity.

Decomposition/solubilization of sludge solids into water soluble components was found to be dependent upon the variables found controlling for the concentration process: settling time and temperature, initial pH, the particular batch of sludge processed. The loss of solubilized solids into the raffinate was also dependent upon the solids content of the feed sludge.

Losses in the raffinates were calculated from a) the TC analysis of the raffinates, centrifuged to remove suspended solids, b) the volumes (weights) of the raffinate, c) the weight of suspended solids in the feed, and d) the TC of the solids. The BOD recycle load to the plant can then be estimated from the correlations established between TC and TOC, and between TOC and BOD_5.

4.4.12 TC Losses Increase with Increasing Settling Time and Temperature

The Esso oil concentration process requires a settling step at temperatures ranging up to ~80°C for times up to ~24 hours. Some thermal decomposition of the sludge solids was therefore expected. That heat treatment at high temperature can result in large losses has been well documented in the literature for known commercial processes (30-33); BOD levels in the recycle stream can be as high as 5000 mg/lit.

As expected for a thermal process, the TC losses with the Esso oil concentration process increase with increasing settling time and settling temperature; the effects for representative runs for several different sludges are shown in Figures 4 and 5 respectively; complete test data are tabulated in Appendices A-20 and A-21. The rate of TC loss with time rapidly decreases for settling times above about 2 hours. Over the temperature range tested, the rate of TC loss into the raffinate appears to increase linearly with temperature up to 60°C and then taper off.

4.4.13 Losses Decrease With Increasing Feed Solids Content

As shown in Figures 6 and 7, for 2 hour and 20 hour settling at 80°C, respectively, raffinate losses were inversely proportional to initial feed solids content; similar results were obtained for lower temperatures as well. The large scatter of the data points around the regression line is believed to be primarily due to the variability in TC losses for the different batches tested. Further work with a few batches at many dilutions would be required to develop a more quantitative correlation curve.

The inverse effect of feed solids content can be explained by the reduction in concentration factor (and corresponding reduction in raffinate volume) with increasing feed solids content, rather than a reduction in rate of solubilization itself.

As previously shown, final solids content after concentration was not effected by initial feed solids content. Since TC losses are inversely related to feed solids content, there is a definite incentive to operate at maximum solids content consistent with overall process economics, considering costs for thickening and heat balance.

4.4.14. Total Carbon Increase in Raffinate not Due to Presence of Oil

One possible source for at least part of the TC in the raffinate was initially considered to be the oil itself; since oil solubilities in water are low, dispersion of oil into very fine droplets due to the high shear mixing seemed possible. Tests with two oils, #4 Heating Oil and #1 Varsol, mixed with the filtered supernate from the sludge showed only small increase in TC after 20 hrs at 80°C compared with normal sludge tests:

FIGURE 4

FEED CARBON IN RAFFINATE
INCREASES WITH SETTLING TIME

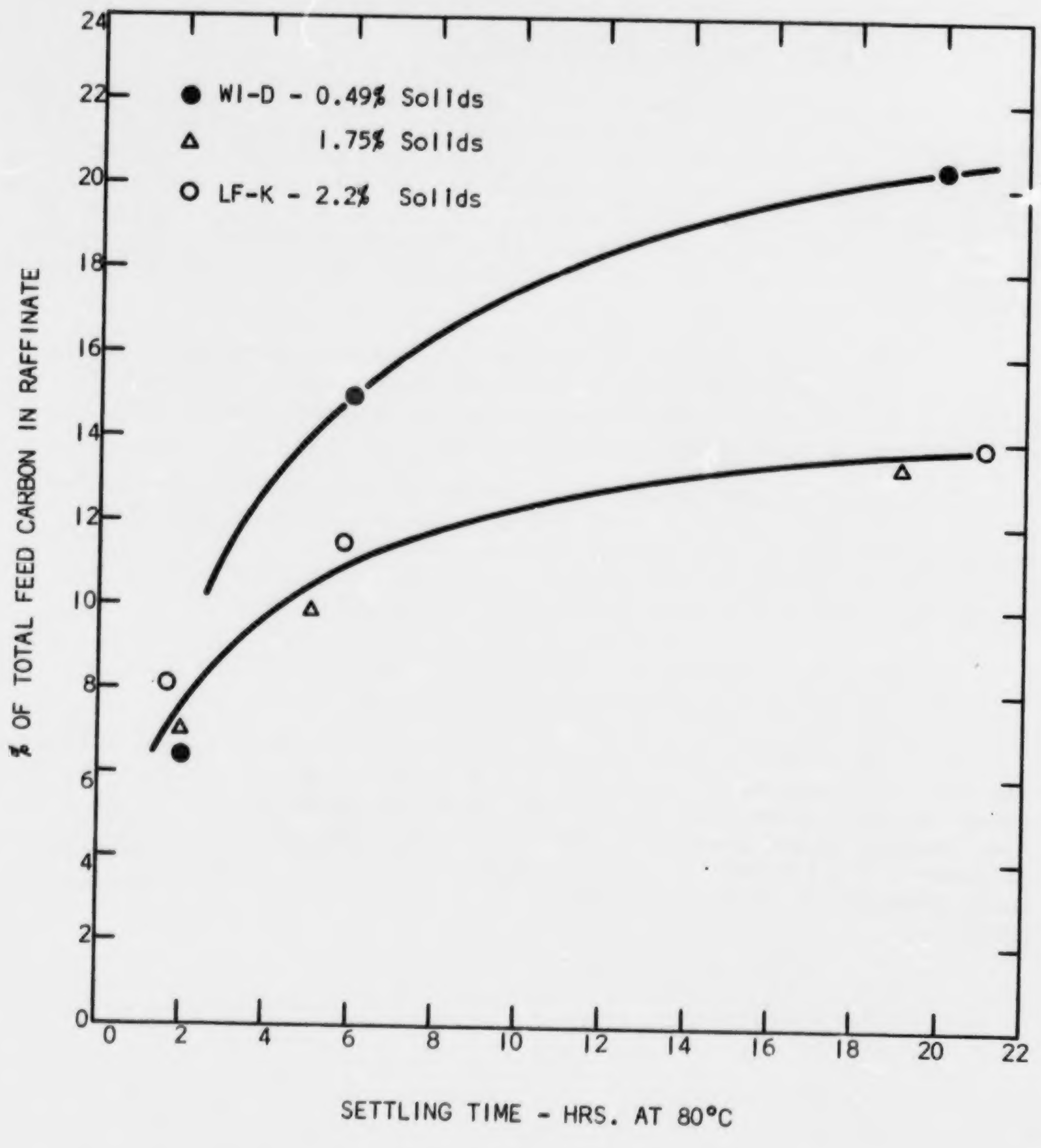
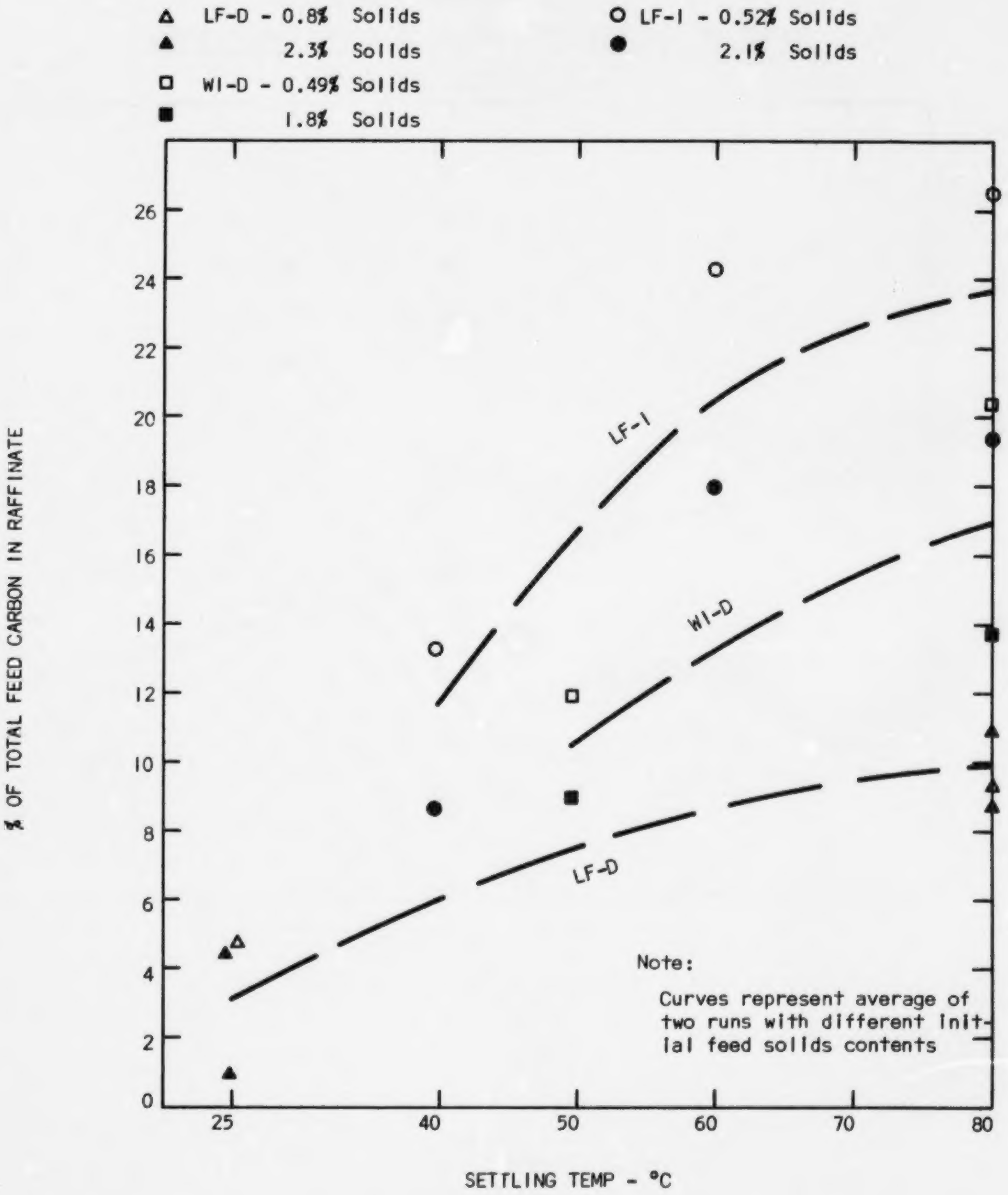


FIGURE 5

TC LOSSES IN RAFFINATE VS SETTLING TEMPERATURE

18/22 Hrs of Settling



Note:

Curves represent average of two runs with different initial feed solids contents

FIGURE 6

EFFECT OF FEED SOLIDS CONTENT ON TOTAL CARBON LOSSES

80°C Settling for 2 Hrs

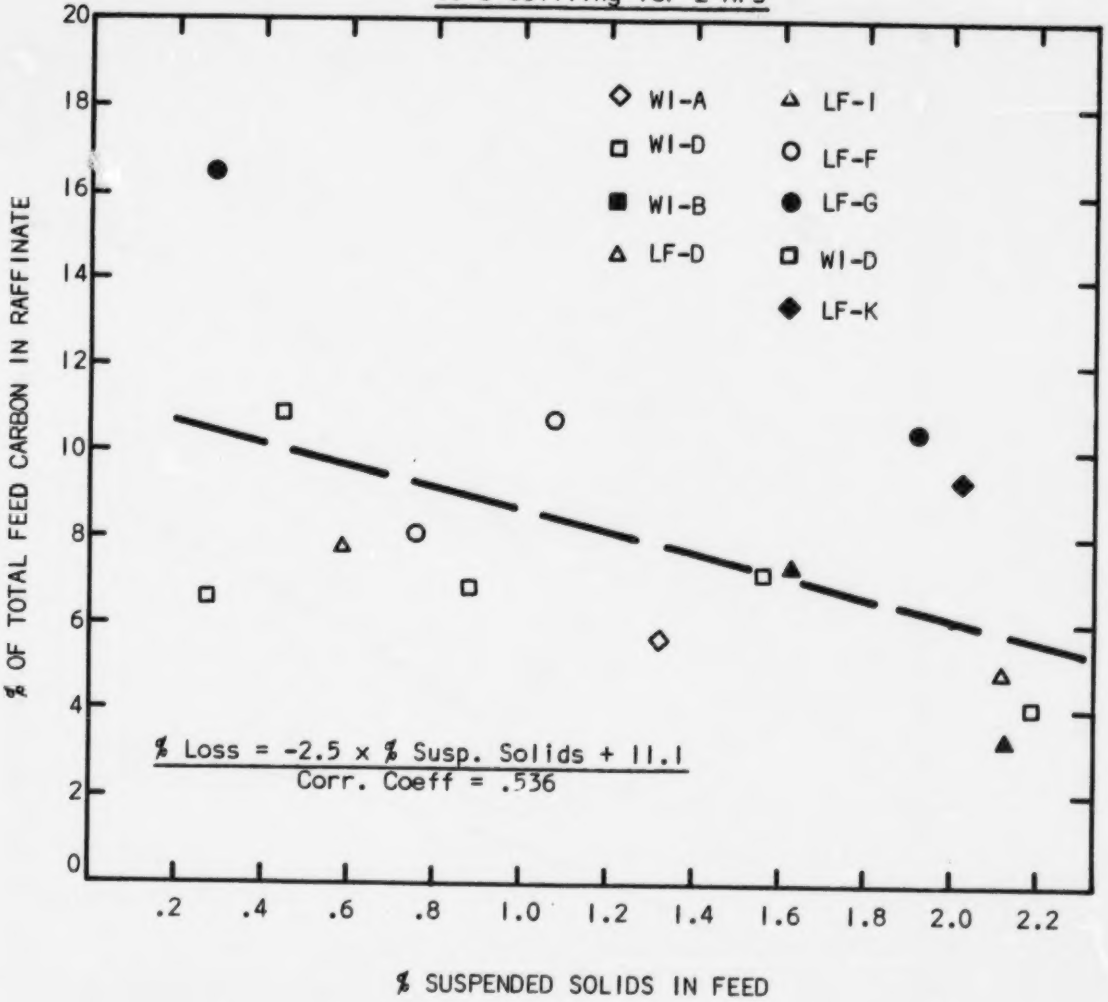


FIGURE 7

EFFECT OF FEED SOLIDS CONTENT ON TC LOSS

80°C Settling for 18/22 Hrs

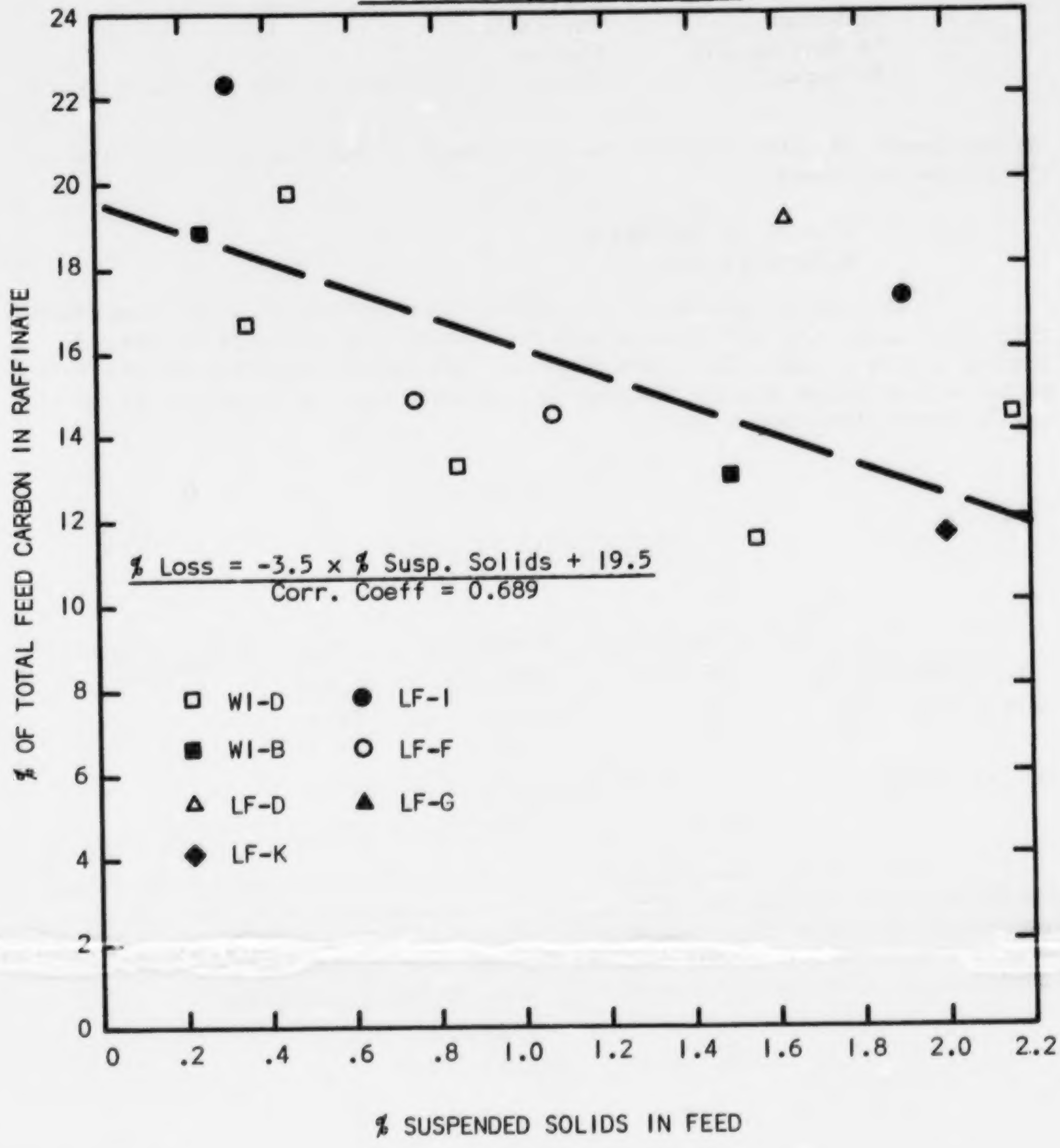


TABLE 14

TC IN RAFFINATE NOT DUE TO OIL

Oil	Mixed With	Increase in TC in Raffinate, ppm
#4 Heating Oil	Supernate	43
#1 Varsol	Supernate	15
#4 Heating Oil	Sludge	1020
#1 Varsol	Sludge	860

On the basis of these results the oil itself is not the source of the TC in the raffinate.

4.4.15 TC Loss in Raffinate Reduced by Oil

As part of the study on factors controlling TC loss, comparison runs were made with and without oil, but using the same conditions of mixing and settling. The percentage of feed solids solubilized was found to be almost twice as high without oil as with oil, as shown in Table 15 for 20 hours settling at 80°C:

TABLE 15

OIL REDUCES TC LOSS

Sludge Batch	% Solids in Feed	Settling Temp. °C	% Feed C Solubilized	
			No Oil	With Oil
Bergen County D	0.8	80	18.4	11.1
	2.3	80	19.7	12.1
Bergen County G	0.55	80	48.7	20
		Average	29	15

Explaining these results without considerably more work is not possible; some form of "shielding" of the sludge solids by the oil is indicated, possibly from the effects of the mixing. This hypothesis is apparently supported by the results of one test comparing agitation vs. no agitation at 25°C, both without oil, with 20 hours settling:

Agitation	% Feed C Solubilized
none	7.3
pump	15.3

The higher solubilization with agitation indicates that cell walls have been ruptured, releasing water soluble compounds.

4.4.16 No Apparent Effect of Oil Type

Comparison of TC losses for #4 Heating Oil and #1 Varsol show no consistent effect of oil. As shown in Table 16, the average TC loss for all tests at 20 hour settling is the same for both oils, with some indication of lower TC with #1 Varsol for short settling time. The lack of an oil effect is also supported by the single tests with #2 Heating Oil and LOPS.

One possibility remains to be tested in regard to effect of oil on TC loss: that the "light ends" in the different oils actually used anesthesize the living bacteria cells, causing leakage of amino acids. Evaluation of a high boiling, lube oil base, such as Coray 37, would be required.

4.4.17 Lowering Sludge pH to 3.0 Reduces TC Loss

In addition to increasing concentration factor, reducing pH also has the desirable effect of reducing TC loss in the raffinate. For the three runs evaluated (see Table 17), the TC loss was 13-24% less than the unadjusted controls, with the average reduction 18%.

The actual reduction obtained in feed solubilized at pH 3 is somewhat greater than the effect above for raffinate loss. At pH 3 the concentration factor is increased by ~15%, so the raffinate volume is correspondingly 4-5% higher than for unadjusted sludge.

4.4.18 Estimate of Recycle Load vs. TC Loss in Raffinate

In calculating the impact of TC losses in the raffinate the most important value is the recycle load to the plant, rather than the TC loss per se. The quantitative translation of TC loss to recycle load depends upon many factors which are specific to each particular plant. A preliminary estimate of the recycle load for a given TC loss level can be made for activated sludge assuming the following:

1. BOD_5 of plant influent = 200 mg/liter
2. BOD_5 of primary effluent = 130 mg/liter
3. Production of waste activated sludge = 655 pounds total solids per MGD influent
4. % Carbon in waste sludge solids = 35%
5. TC/TOC ratio for the raffinate recycled = 1.17
6. BOD_5/TOC ratio for the raffinate recycled = 2/1

EFFECT OF OIL ON TC LOSS IN RAFFINATE

Sludge Batch	% Suspended Solids in Feed	Settling		% Feed C in in Raffinate			
		Temp. °C	Time-hrs	#4HO	#1 Varsol	#2HO	LOPS
WI - A	1.5	80(1)	3	5.8	3.1	5.6	
	1.5		20	9.5			12.9
B	2.35	80(1)	1	1.9	2.1		
	1.65		7	6.0	5.0		
	1.65		20	11	12		
	0.55		20	18.8	18.8		
LF - D	2.3	25(1)	1	1.7	0.2		
	2.3		22	4.3	0.9		
	2.3	80(1)	1	3.9	2.7		
	2.3		20	10.8	8.8		
LF - F	.95	70(1)	18	16.7			
	1.28		20		16.4		
WI - D	0.66	70(1)	21	21.7			
	1.06		24		22.2		
Averages:		70/80(2)	1/7	3.8	2.6		
		70/80(2)	18/24	15.8	15.6		

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(1) Laboratory runs

(2) Pilot Plant runs

TABLE 17

TC LOSS REDUCED WITH SLUDGE AT pH 3

Sludge Batch	% Susp. Solids in Feed	Settling Temp. °C	Settling Time Hrs.	TC Loss in Raffinate, % pH ~6.5 (2)	TC Loss in Raffinate, % pH 3.0	Ratio of TC Loss pH 3/7
Wards Island	0.49	80	2	6.5	10.4	
(W.I. "D")			6	15.8	16.4	
			20	21.0	18.2	.87
	1.75	80	2	7.0	5.3	
			20	13.4	10.1	.76
Bergen County	2.2	40	2	4.8	4.8	
(L.F. "K")			5.5	6.4	5.7	
		80 (1)	7.5	8.3	7.0	
			21	11.5	9.7	.84

(1) Settling temperature increased after 5.5 hrs. ("staged" temperature).

(2) Sludge as received from sewage plant.

The first three assumptions were recommended by the EPA. Item 4 was based on analysis of sludge solids tested in this program (see Table 2), items 5 and 6 on analytical data obtained during the pilot plant program (Phase 3).

Based on these assumptions, a 10% loss of sludge solids, based on TC loss in the raffinate would be equivalent to about 3.7% of the BOD_5 load to the secondary treatment, or about 2.5% of the BOD influent to the plant.

4.4.19 TC Losses in Raffinate Dependent Upon Sludge Batch

Review of all data on TC losses clearly shows the large variability between batches for a constant set of test conditions. As summarized in Table 18, TC losses for 20 hours at 80°C, with feed solids content of 1.5%, varied from 6.5-24% for the different types of sludge; the range for one type of sludge, Bergen County activated, was 10-24%.

TABLE 18

TC LOSSES IN RAFFINATE DEPENDENT UPON SLUDGE BATCH

Sludge Batch		% Feed C in Raffinate (1)
Bergen County	D	10
	F	16
	G	23
	I	24
	K	17
Wards Island	A	11
	B	15
	D	16
Trenton	B	6.5

---

(1) All values adjusted to 20 hrs at 80°C, 1.5% solids in feed sludge.

At this time no information is available which provides an explanation for the variability shown above. Since the test conditions were, to the best of our knowledge, closely controlled the variability is assumed to reflect differences in chemical composition and physical structure.

4.4.20 Batch/Batch and Plant/Plant Variability in Solids Concentration Achieved

Using the data available from the laboratory program, some preliminary indication of variability in response to the concentration process can be obtained; the differences of interest are batch/batch for a given sludge type and source (plant) and plant/plant. This comparison is summarized below, using results obtained with high shear mixing only, and for 20/22 hours settling at 80°C:

TABLE 19

VARIABILITY IN FINAL SOLIDS CONCENTRATION

Sludge Source	Type	Number of Batches Compared	Range of % Solids in Feed	Range of % Solids in Concentration (1)
Bergen County	activated	5	0.5-2.1	7.5-10
Bergen County	primary + activated	2	1.7-3.5	6-12
Wards Island	activated	3	0.5-1.8	6-8
	primary + activated	1	2.1	11
Trenton	trickling filter	1	1.2	7

(1) Not corrected for % oil in sludge or TC losses in raffinate.

The variability between batches of the same sludge type and source, as well as between plants for the same sludge type, is considerable. This variability is believed primarily due to intrinsic differences, rather than experimental error or test reproducibility. The large differences in sludge thickening properties support the assumption of intrinsic differences in the sludge batches and the variable response to the oil concentration process. The differences between the two batches of Bergen County primary + secondary may reflect sampling problems as well as inherent variability.

Several batches, from each of many more plants, would have to be evaluated for response to the concentration process to establish the expected range of variability of the process in general commercial usage.

4.4.21 Staged Settling Offers Advantages

Previous data on TC loss vs. settling temperature have clearly established that TC loss increases rapidly with temperature. Since the concentration factor also decreased with decreasing temperature, however, use of low temperature settling might not prove to be economically practical. A staged settling process, where most of the water is removed in the first stage(s) at low temperature, with final concentration at high temperature, was considered a promising approach; the hope was to reduce TC loss without sacrificing solids concentration. The other important advantage of a temperature staged extraction process is the large savings in heat requirement, since less than 1/2 the initial water in the feed sludge will remain for heating to 80°C.

In the actual laboratory tests, the first stage settling was carried out for 5-1/2 hours at 40°C, and second stage at 80°C for an additional 15 hours. The TC loss was 15% lower for the one test run by staging (see Table 17). As shown in Table 20, the final solids concentration achieved by staging was about the same as the straight 80°C reference run at high feed solids content, and considerably lower than the reference at the low feed solids. The difference in response is believed due to the reduced effect of temperature on concentration factor with increasing feed solids content; this directional effect has been previously noted in the laboratory variable study program for constant temperature settling.

The general effectiveness of the temperature staged settling has been qualitatively confirmed in the pilot plant program, (discussed in the section on Phase 3) since these tests were actually run on a staged basis. The feed sludge was at ambient temperature at the start of the run and required several hours (3-4) to reach final settling temperature in the jacketed settler; water raffinate was removed periodically during the run, including this warmup period. While direct, controlled comparisons, such as the above laboratory tests, were not made, the solids concentrations achieved were equal to the values for the isothermal laboratory runs with the same sludge or sludge type; TC losses were not consistently lower, however. Taken together with the laboratory tests, the pilot plant results appear to confirm the feasibility of the temperature-staged settling.

The possibility of utilizing the staged concept for pH adjustment, in order to obtain higher solids concentration at pH 3, was also evaluated on a preliminary basis. The oil sludge concentrate obtained from pH ∿6.5 sludge after "regular" 24 hour settling was adjusted to pH 4 and to 2, then settled for an additional 16 hours. Without pH adjustment, no further increase in concentration was obtained; with pH adjustment, appreciable increases in concentrations were obtained as shown below (on page 46).

TABLE 20

EFFECT OF STAGED TEMPERATURE SETTLING ON SOLIDS CONCENTRATION

Bergen County "K" Activated Sludge

% Susp. Solids in Feed	Settling Temp. °C	Settling Time, Hrs.	% Solids in Concentrate(2) pH ~6.5(3)	pH 3.0
0.5	40	2	3.9	4.2
		21	6.5	6.8
	80	2	5.9	
		21	9.2	
	40-80(1)	2	4.5	
		21	7.4	
2.2	40	2	5.3	
		21	8.2	
	80	2	6.6	
		21	9.7	
	40-80(1)	2	4.8	5.7
		21	9.9	12.2

(1) Settling temperature increased to 80°C after 5 hrs.

(2) Not corrected for TC losses.

(3) Sludge as received.

Sludge pH	Increase in Solids Concentration, % Test 1	Test 2
6.4	0	0
4.0	11	
2.0	22	27

Since stainless steel equipment is required for operation below pH 5.5, staging can be used to minimize costs: with staged settling only the final concentration stage will require corrosion resistant equipment.

4.4.22 Oil Recycle Not Detrimental to Concentration Step

One of the major concerns early in the program was the effect of oil recycle on the efficiency of the extraction process; specifically, would surface active compounds, formed by thermal degradation, build up in the oil and gradually reduce the concentration factor obtained. Three separate tests were carried out in connection with this problem, evaluating a) recycle oil from the Hershey, Pa. plant using the Carver-Greenfield process to dry mixed primary + secondary sludge, b) recycle oil from the Carver Greenfield pilot plant test (Phase 3), c) simulated recycle oil prepared in the laboratory.

The recycle oil from Hershey, which had been through an undefined number of cycles, contained several percent of calcium stearates, plus an appreciable amount of fatty acids and nitrogen containing compounds (see Appendix A-22). This oil from the Hershey operation is Coray 37, an unrefined lube oil base stock. The oil from Carver Greenfield, which was processed once through the entire concentration-evaporation cycle, was #4 Heating Oil. Laboratory simulation tests were carried out by refluxing sludge with the recycle oil from Carver Greenfield for several hours, then centrifuging to recover the oil; this procedure was repeated on a portion of the oil for greater severity.

The effects of oil recycle, using the oils described above, were evaluated for several different batches of Bergen County activated sludge, and of several different feed solids contents. Test data, tabulated in Table 21 and summarized below, show essentially no difference between fresh vs. recycle oil:

Type Oil	Concentration Factors (18/20 hrs at 80°C) 1 Hr Settling	20 Hr Settling
Fresh oil	5.6	8.9
Recycle oil	5.6	8.7

Based on the data from these seven tests, there is no reason for concern about adverse effects with recycle oil.

TABLE 21

EFFECT OF OIL RECYCLE ON EXTRACTION

Bergen County - Activated Sludge
All Runs at Oil/Sludge ratio of 0.2, 80°C Settling

Test(6)	% Susp. Solids in Feeds	Type Oil	% Solids in Concentrate(1) 0.8/1.5 hrs	5.5 hrs	18/20 hrs
1	0.84	Fresh Coray 37	2.7		6.3
	2.3	recycle "	3.1		6.6
2	1.72	Fresh #4 H.O.	4.1		7.9
		1 recycle(2) "	3.6		6.6
3	0.75	Fresh "	4.2		9.2
		1 recycle(2) "	4.1		8.5
	1.5	Fresh "	5.4		10.3
		1 recycle(2) "	5.3		9.3
	3.0	Fresh "	5.3		10.1
		1 recycle(2) "	4.9		10.0
4	1.0	Fresh "		8.0(4)	8.9(4)
		multiple recycles(3)		8.2(4)	9.4(4)
5	0.50	Fresh "		6.6	8.2
		multiple recycles(5)		7.6	9.0
6	2.2	Fresh "		8.1	9.9
		multiple recycles(5)		8.4	10.2
		Average fresh	5.6		8.9
		Average recycle	5.6		8.7

---

(1) Uncorrected for TC losses in raffinate.
(2) Processed once thru Carver Greenfield concentration step.
(3) Oil from (2) refluxed for 2 hours with sludge, then centrifuged.
(4) Average of 2 runs.
(5) Oil-sludge from (3) refluxed for 4 hrs, then centrifuged.
(6) Test 1 batch LF-"D", Tests 2-5 batch LF-"J", test 6 batch LF-"K".

4.4.23 Concentration of Sludge Without Oil

Control runs without oil, but employing the standard agitation used for the oil process, have been made for most sludge batches processed in the laboratory. For most of the batches the sludge solids were concentrated by "floating" to the top of the settling vessel. As shown in the data summary in Table 22, however, there is no apparent pattern or consistency to the occurrence of this "no-oil" type concentration in terms of initial solids content, type of agitation, or settling temperature. This lack of consistency in response (settling or floating) occurred with different batches from the same sludge source.

In the cases where the solids floated, gas bubbles at the bottom and mixed in with the sludge concentrate phase were observed. The solids did not float unless agitated first. The interface between the concentrated sludge and the relatively clear liquid was much less stable than when the oil concentration process was used; even slight disturbance of the settler caused some of the solids in the "float" layer to detach and settle out in the raffinate.

As discussed in the section on TC losses in the raffinate, the losses were more than twice as high for the control runs without oil as with the oil present; this relative difference occurred at both 25°C and 80°C.

Considering the unpredictability of the concentration achieved without oil, the higher TC losses incurred, and the unstable interface, there was no apparent incentive to consider this approach further.

TABLE 22

SLUDGE CONCENTRATION WITHOUT OIL

Sludge Batch (1)	Sludge Type	% Susp. Solids	Storage (2) Time, Days	Agitation	Type Sludge Response	Settling Temp.	Comment
L.F.-"B"	activated	0.9	6	5 sec W. Blender	solids settled	22°C	comparable settling to no agitation
"	"	1.7	6	30 sec propeller	" "	"	
L.F.-"E"	"	1.0	1	5 sec W. Blender	" "	25°C	slower settling than no agitation
				20 sec turbine	" "	"	
				2 x thru pump	" "	"	
W.I.-"B"	"	0.55	6	2 x thru pump	" "	25°C	slower settling than no agitation
W.I.-"D"	primary + secondary	2.1	10	1 x thru pump	solids floated	80°C(3)	comparable conc. to oil extraction
W.I.-"D"	activated	0.6	1	1 x thru pump	solids settled	25°C	
				"	split; mostly settled same floated	60°C	
L.F.-"G"	activated	0.4	15	1 x thru pump	solids settled	25°C	very little con-concentration; C.F= 1.2 in 22 hrs.
		1.8		"	solids floated	"	
L.F.-"H"	secondary	0.53	12	1 x thru pump	solids floated	25°C	
				"	" "	60°C	
L.F. "I"	activated		1/2	1 x thru pump	" "	50°C	
	"		"	"	" "	80°C	
	"		"	"	" "	25°C	
L.F. "I"	primary + secondary		"	1 x thru pump	" "	50°C(2)	
	"		"	"	" "	80°C(2)	
Trenton B	Trickle Filter		1	1 x thru pump	solids settled	25°C	
					" split	60°C	

(1) L.F. = Bergen County, W.I. = Wards Island.
(2) At 40°F in refrigerator.
(3) Separation started immediately after mixing and before sample heated up; so same result can be assumed at 25°C.

5. PHASE 2: PILOT PLANT SCALE UP

After completion of the laboratory process and optimization studies, a pilot plant program was then carried out, with two broad objectives:

- Produce the oil-sludge concentrates required for the heat transfer studies of Carver-Greenfield.
- Confirm the small scale laboratory results and determine if there were any unexpected scale-up problems.

The program was set up on the basis of a total of seven runs. These runs were selected to study a) a range of sludge sources and types, and b) oil types considered necessary for the heat transfer program. Oil viscosity was considered to be an important consideration by Carver-Greenfield; therefore, #4 heating oil and #1 Varsol were selected to provide a wide practical range for comparison.

The process parameters of concern for the pilot plant program, in terms of scale up from the 400 cc laboratory scale to the 200 gallon pilot plant scale, were a) concentration factor achieved, b) total carbon losses in the raffinate, c) suitability of the centrifugal pump for mixing of the oil and sludge.

5.1 Description of Pilot Plant Operation

The design basis for the pilot plant was patterned very closely on the laboratory operation: batch premixing of the oil and sludge, process mixing with a centrifugal pump to impart high shear contacting, and finally batch settling to produce the oil sludge concentrate. Detailed operating procedures and description of the equipment are given in Appendix B-1. Setting up and operating a continuous pilot plant for all steps, while desirable from a process demonstration aspect and to provide additional data for detailed design of a commercial plant, was beyond the scope of the current project in terms of manpower and cost.

Mechanically the system worked well, with only two problems developing at the end of the program:

- Plugging of the sludge transfer pump and/or the sludge tanks discharge line. This problem was caused by the fibrous solids in the mixed primary and secondary sludge and will not be encountered in commercial-size operation, employing larger pumps and lines and a solids grinder (comminuter) ahead of the concentration process.

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- Excessive compaction of sludge solids after batch settling in the sludge storage tank; channeling developed in the compacted solids, which resulted in high supernant and low solids content in the sludge removed for a run. This would not occur in a commercial thickener with continuous flow.

For all runs, sludge at ambient temperature (65-75°F) was mixed with 250°F oil and pumped by a centrifugal pump "mixer" to the jacketed batch settler; the jacket temperature was set at 175°F for the initial runs and 185°F for the later runs. Final concentrate temperatures under these conditions were 158°-165°F; higher jacket temperatures were not used to avoid excessive overheating of the material along the settler wall. Since the heat transfer surface/batch volume ratio was low due to settler configuration, 2-4 hours were required for the average batch temperature to reach steady state value.

The test program, in terms of sludge and oil types selected, and initial feed solids contents actually processed are summarized below:

TABLE 23

PILOT PLANT TEST PROGRAM

Run No.	Sludge Feed		Oil	% Suspended Solids in Feed
	Source	Type		
1	Bergen County	Activated	#4 Heating Oil	0.9
2	" "	"	#1 Varsol	1.3
3	" "	"	#4 Heating Oil	1.8
4	Wards Island	"	" " "	0.6
5	" "	"	#1 Varsol	1.1
6	Bergen County	Mixed Primary + Activated	#4 Heating Oil	1.7
7	Trenton	Trickle Filter	" " "	1.6

Results of the pilot plant program, in terms of concentration factors achieved and total carbon losses in the raffinate, are summarized in Table 24 and discussed in detail in the sections below.

5.2 Pilot Plant Scale-Up Correlates Well With Laboratory Results

5.2.1 Solids Content After Concentration

One of the objectives of the pilot plant operation was to determine the effect of scale-up, if any, on concentration factor. Comparison of pilot plant and laboratory results shows satisfactory agreement on concentrations achieved, with the pilot plant solids contents at least equal to the

TABLE 24

SUMMARY OF OPERATING DATA FOR PILOT PLANT RUNS

Run	Sludge Source	Sludge Batch	Type	% Feed Solids	Oil Used	Hrs. Settling	Settling Temp. °C(1)	Concentration Factor	TC Loss in Raffinate-% (2)
1	Bergen County	F	Activated	0.95	#4 HO	1	50	1.0	--
						18	70	7.8	16.7
2	Bergen County	F	Activated	1.28	#1 Varsol	1	50	1.0	0.2
						2	66	2.7	11.7
						20	70	5.6	16.4
3	Bergen County	G	Activated	1.82	#4 HO	1	40	1.0	0.3
						3	70	1.6	8.7
						20	70	4.8	21.0
4	Wards Island	D	Activated	0.66	#4 HO	1	60	3.9	
						2	70	4.7	8.9
						4	70	5.5	11.3
						21	68	10.4	21.7
5	Wards Island	D	Activated	1.06	#1 Varsol	1	60	1.1	0.4
						4	60	3.6	8.8
						7	67	4.7	11.8
						24	72	6.8	15.1
6	Bergen County	I	Primary + Activated	1.68	#4 HO	1	62	2.4	5.0
						4	65	3.0	8.1
						21	74	4.7	14.1
						27	74	5.0	15.0
7	Trenton	B	Trickling Filter	1.63	#4 HO	1	55	1.1	0.15
						2	55		
						3	61	1.2	0.36
						16	80	3.5	5.9
						22	80	4.3	6.5

---

(1) Temperature of water raffinate at time of sampling.
(2) % of TC in feed solids.

corresponding laboratory values; for the seven runs, the average solids content of the sludge concentrate, not corrected for oil solubles or TC losses, was ∿7.6% for the pilot plant vs. ∿7.3% for the laboratory (see Table 25).

TABLE 25

COMPARISON OF CONCENTRATE SOLIDS CONTENTS OF PILOT PLANT & LABORATORY RUNS

Run No.	Feed Sludge			Settled Oil Sludge[1] Uncorrected Solids Content - %	
	Type	Source	% Susp. Solids	Pilot Plant	Lab (2)
1	Activated	Bergen County	0.95	6.7	7[3]
2	Activated	Bergen County	1.28	7.7	7.5[3]
3	Activated	Bergen County	1.82	8.7	8[3]
4	Activated	Wards Is.	0.66	6.8	7[4]
5	Activated	Wards Is.	1.06	7.9	7[4]
6	Primary + Activated	Bergen County	1.68	8.8	8[4]
7	Trickling Filter	Trenton	1.63	6.9	6.5[4]

---

(1) 18/22 hrs settling, no corrections for solubilized sludge solids or oil soluble fraction.

(2) Interpolated to match pilot plant conditions.

(3) Composite average of all Bergen County batches.

(4) Results on same batch as processed in pilot plant.

The satisfactory agreement between the solids concentrations obtained in the pilot plant and the equivalent laboratory runs at least directionally confirms the effectiveness of the staged settling technique. In the laboratory test of this technique the oil-sludge mixture was first settled for 5 hours at 40°C (105°F) to remove 50-60% of the water, followed by 15 hours at 80°C (175°F). The pilot plant runs were actually a modified form of staged settling, since the oil sludge mixture after contacting was only about 90°F and required 3-4 hours in the jacketed settler to reach the desired temperature.

In terms of further scale-up of the concentration (settling) step to plant scale, which involves the rate of separation of the oil + sludge and the water raffinate phases, the above results are encouraging. The initial liquid depths at the start of the concentration step were 0.50' & 0.65' in the laboratory tests vs. 1.75' in the pilot plant runs. The relative rate of separation of the oil sludge concentrate phase, which controls the design of the settler, apparently was not effected by the 3.5/1 increase in liquid depth. The liquid depth of 1.75' for the oil-sludge mixture at the start of the settling step was used for the design of plant size settlers, recognizing that this represented a very conservative design basis. A further substantial increase in initial liquid depth (to 3-4') without adversely effecting settling rate could reasonably be assumed on the basis of the scale-up results, but cannot be used with confidence until actual experimental verification is obtained. As will be discussed further in Phase 4, confirmation of an increased scale-up factor would permit a substantial reduction in settler cost.

5.2.2 Centrifugal Pump Satisfactory Mixer

The degree of solids capture with a centrifugal pump was excellent in the laboratory, with no reduction in effectiveness at low sludge suspended solids contents. Since scale up of mixing effects with a centrifugal pump is not well defined, one of the objectives of the pilot plant program was to confirm the laboratory results showing high solids capture. Quantitative analysis of the 3 runs confirmed this observation:

Run	% Suspended Solids in Feed	Solids Capture %
3	1.82	97.3
4	0.66	98.3
5	1.06	97.2

These values for solids capture are, if anything, too low; a small portion of the feed solids was not contacted by the oil in the mixing step, but was trapped in the lines during the sludge transfer step and charged directly to the settler.

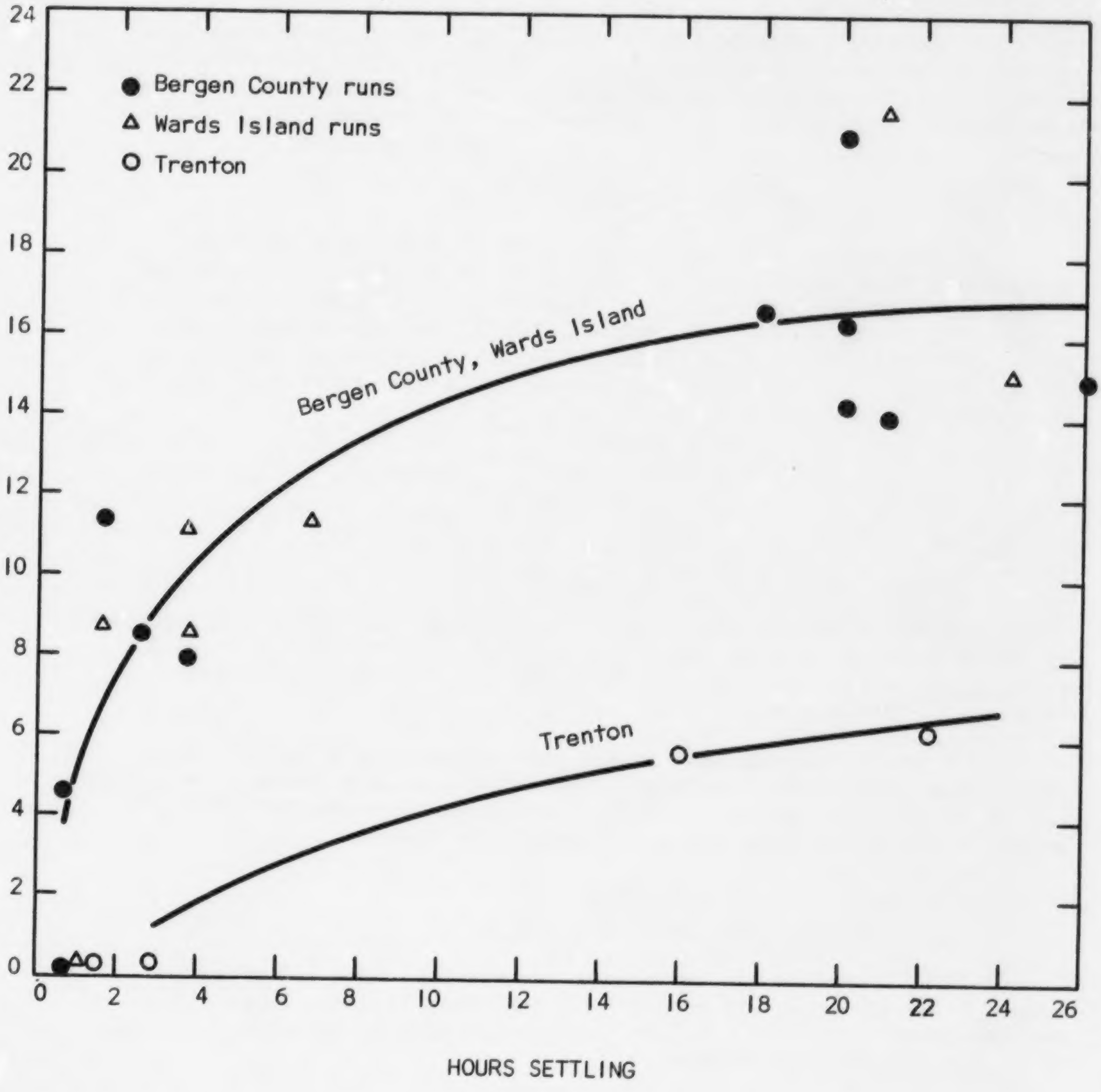
The contacting time in the centrifugal pump mixer was adjusted to match that used in the laboratory runs. The agreement between laboratory and pilot plant concentration factors and solids capture indicates that satisfactory scale-up of mixing intensity was obtained.

5.2.3 Total Carbon (TC) Losses Are Consistent with Laboratory Data

TC losses in the raffinates vs. settling time are shown in Figure 8 for all runs; the curves have the characteristic shape found for the laboratory studies on TC loss, and are grouped within a fairly close range for similar type sludges.

FIGURE 8

TOTAL CARBON LOSSES IN RAFFINATE - PILOT PLANT RUNS



Losses for 18/21 hours settling were in the range of 15-22% for the Bergen County and Wards Island activated sludge vs. 6.5% for the Trenton trickling filter sludge. Without prior and more extensive data on trickling filter sludge, we can only assume either that a) the rate of solubilization/decomposition for this type is markedly different from activated sludge or b) this low value reflects the large batch/batch variability noted in the laboratory program for TC losses in the raffinate.

As shown in Figure 9, the TC losses for the pilot plant runs are consistent with these values obtained in the laboratory program. Considering the range of values found for TC losses for different sludge batches, the pilot plant results can be considered as confirming the laboratory results. Since TC losses are a function primarily of kinetic parameters (temperature, time) not involving scale-up factors, good agreement between laboratory and pilot plant was expected.

The agreement between laboratory and pilot plant for mixing, concentration factor and TC loss discussed above is important for two reasons: confidence is established in the feasibility of increasing the scale-up factor to commercial size, and results of any laboratory studies can be extrapolated to commercial scale as required for the process trade-off studies in Phase 4.

5.3 Analysis of Raffinates

5.3.1 Soluble Organics in Raffinate are Biodegradable

One of the concerns about the TOC in the raffinate was the BOD equivalent, which is important in calculating recycle load and effect on the overall treatment plant. As shown in Figure 10, the BOD_5/TOC correlation factor for two pilot plant runs ranges from 1.67-2.47/1 over the range tested (TOC range 60-1620 ppm); this is consistent with the literature correlation for BOD vs TOC in effluent streams from conventional processes at the lower TC levels (6). The recycled organics from the Esso process will therefore be normally biodegradable in a secondary treatment plant.

In the initial evaluations of raffinate losses, both laboratory and pilot plant, TOC analysis was obtained in addition to TC. A consistent correlation between TC and TOC of about 1.1-1.25 was found with an average of 1.16 so the dual analysis system was dropped and only TC analysis obtained. All raffinate losses are therefore calculated on the basis of TC; where conversion to BOD_5 equivalent is needed, a TC/TOC factor of 1.16 was used, along with a BOD/TOC^5 factor of 2.0. The limited data on BOD/TOC ratio show a fairly wide variation for the TOC range evaluated; further comparisons at the high TOC level would be required to establish a more precise correlation.

FIGURE 9

PILOT PLANT TOTAL CARBON LOSSES
CONSISTENT WITH LABORATORY DATA

18/22 HOURS SETTLING

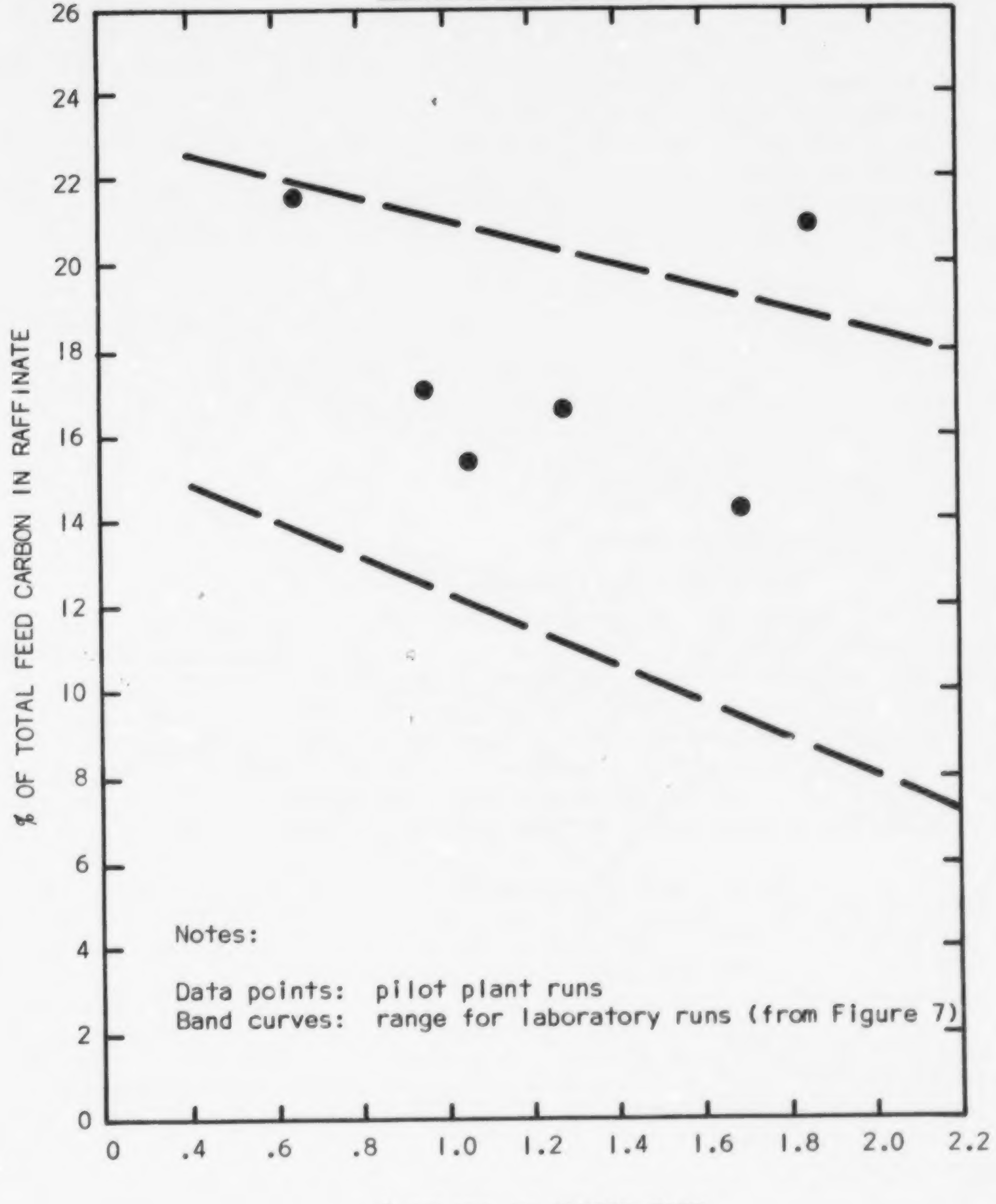
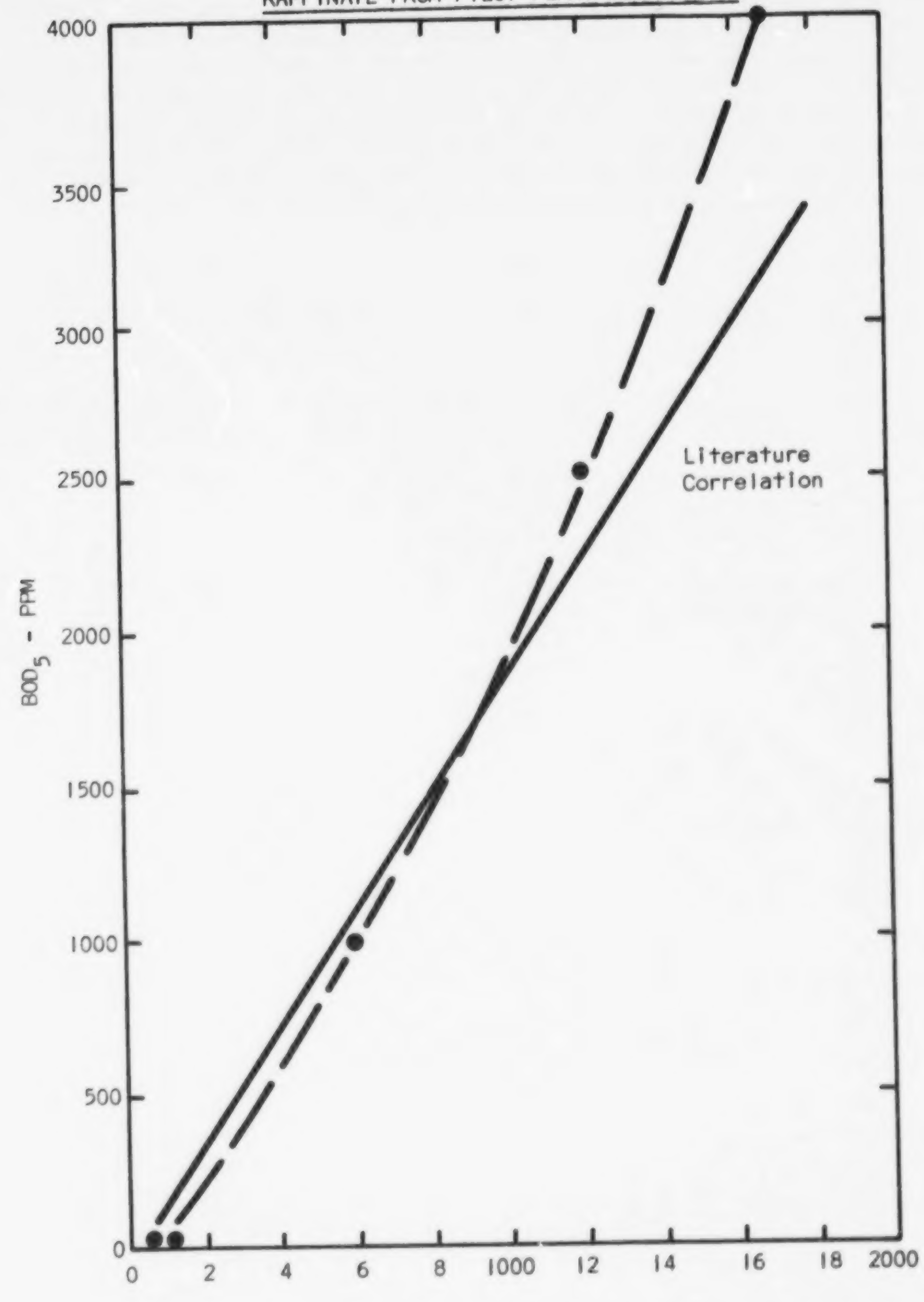


FIGURE 10

BOD5 VS TOC

RAFFINATE FROM PILOT PLANT RUNS 1 & 3



TOTAL ORGANIC CARBON (TOC) - PPM

5.3.2 Nitrogen and Phosphorous

To obtain some indication of the effect of the oil concentration process on N and P in the sludge solids, initial sludge supernatants and the raffinates were analyzed for NH_4 and NO_3 nitrogen and for phosphorus. The available results detailed in Table 26 show a 400-500% increase in ammonium N, almost a 50% reduction in nitrate N, and about a 50% increase in phosphorus. The large increase in ammoniacal N is assumed to reflect the solubilization/decomposition of the proteins and amino compounds in the sludge solids. The decrease in nitrate suggests some form of denitrification. Further interpretation and explanation of these analyses is beyond the scope of this project, however.

TABLE 26

RAFFINATE ANALYSES - PILOT PLANT RUNS

Run No.	Sludge Type and Source	Feed Supernate - ppm			Raffinate - ppm		
		NH_4-N	NO_3-N	P_2O_5	NH_4-N	NO_3N	P_2O_5
1	Bergen County Activated	29	30	38	133	12	60
3	"						
4	Wards Island Activated	42	42	42	210	20	60
5	"				190	24	82

6. PHASE 3: DETERMINATION OF HEAT TRANSFER PROPERTIES

6.1 Basis for Carver-Greenfield Test Program

What is referred to as the "Carver-Greenfield" component of the total process includes the following principal elements:

- A multiple effect evaporator system to remove essentially all of the water from the oil-sludge concentration produced in the Basco concentration.
- Solid bowl centrifuge to separate most of the oil from the dried sludge solids, for recycle to the sludge concentration step.
- Solvent stripping of the solids to remove additional oil in excess of that required for heat balance.
- Incineration of the sludge solids and remaining oil required to a) reduce the solids to inorganic ash which permits disposal with minimum pollution problems, b) generate the heat needed for the evaporation step.

The investment cost of the multiple effect evaporator system depends upon the number of effects required and the size of each effect. The number of effects is determined primarily by heat balance considerations, i.e. the pounds of steam required to evaporate a pound of water in the feed stream vs. the pounds of steam generated by the incineration of the sludge solids and associated oil. By increasing the number of effects the heat balance becomes more favorable. As a generalization, an evaporation system is designed to optimize this balance between investment (number of effects and size of boiler to produce the steam) vs. cost of added fuel oil. The incinerator-boiler cost is primarily a function of the total quantity of steam generated, which in turn is dependent upon the evaporation load (total pounds of water to be evaporated) and the number of effects (which fixes pounds of steam required to evaporate one pound of water).

The size of each evaporator is determined by the total evaporation load and by the rate at which heat can be transferred from the condensing steam to the evaporating process stream. This overall heat transfer rate, in turn, is controlled by a) the operating temperature differential between heat source and heat sink and b) the overall heat transfer coefficient (U). The temperature differential for each effect is, in part, dependent on the design of the evaporator system, but generally must be held within fairly narrow limits by both operating and economic design requirements. For a given temperature differential, then, the controlling factor becomes the overall heat transfer coefficient which is determined primarily by the nature of the process stream being evaporated.

Streams that are fluid, with low viscosities, normally have much less resistance to heat flow than viscous fluids for the same flow rate through the evaporators. The viscosity and associated flow properties are very much dependent upon the physical-chemical characteristics of the process stream. As the oil sludge mixture is concentrated in the first and second effects, the viscosity increases and the overall heat transfer coefficient can therefore decrease. The specific type and source of sludge, as well as the oil used as the fluidizing medium in the evaporation, also can effect the overall heat transfer coefficient.

In a conventional evaporation system, where the process stream consists of an aqueous phase only, complete drying of the sludge solids would be almost physically impossible. At a point well before dryness the straight sludge concentrate would be too viscous to be circulated for further evaporation; in addition, considerable scaling of the heat transfer surface would occur due to solids deposition, drastically lowering the U value. The Carver-Greenfield process minimizes the increase in viscosity and the extent of scaling by utilizing a water insoluble oil as a fluidizing carrier for the sludge solids. Thus, though water is being completely removed the process stream is still kept fluid, even in the last effect (drying stage), and heat transfer rates are kept high. This is the key to the success of the patented, commercially demonstrated, Carver-Greenfield evaporation process (51).

Since the investment cost of the Carver-Greenfield process is a large fraction of the total cost, it must be estimated as accurately as possible within the limits of the project. The specific objectives of the test program carried out by the Carver-Greenfield Co. were therefore set as follows:

- Determine the overall heat transfer coefficients for the oil sludge concentrates from the Esso concentration process, as a function of sludge type and source, feed solids concentration, oil type.
- Evaluate the de-oiling characteristics of the solids after the drying stage, to be able to predict the oil recovery for different sludge types and oils used.
- Determine the organic contaminant levels of the distillate fractions as a function of process conditions (both Esso and Carver-Greenfield process components).
- Provide the necessary data for and assist Esso in the cost optimization of the Carver-Greenfield evaporation process, and in the preparation of a prototype design for a plant.

6.2 Evaluation of Heat Transfer Coefficients in Carver-Greenfield Pilot Plant

The seven oil-sludge concentrates prepared by Esso (see Phase 3, Pilot Plant Program) were evaluated according to the program above in the Carver-Greenfield pilot plant, located in Hanover, N.J. For the tests, the oil-sludge concentrates were remixed in a batch feed tank, then fed to a single effect recirculating evaporator; the partially concentrated effluent was collected and recycled thru the evaporator as required to simulate operation in a multiple effect system. Vacuum and steam temperatures were adjusted as required for simulation of the particular effect being tested. A schematic flow diagram of the Carver Greenfield pilot plant is shown in Figure 11.

The heat transfer resistance of a test oil-sludge process stream in the evaporator is defined by means of the following equation:

$$U = \frac{q}{A(T_s - T)} = \frac{W_s H_s}{A(T_s - T)}$$

where q = rate of heat transfer in Btu/hr

A = area of heating surface ft^2

U = overall heat transfer coefficient for the system

= to the reciprocal of the overall system heat transfer resistance BTU/hr-ft^2-°F

T_s = condensation temperature of the steam °F

T = boiling temperature of the oil-sludge mixture °F

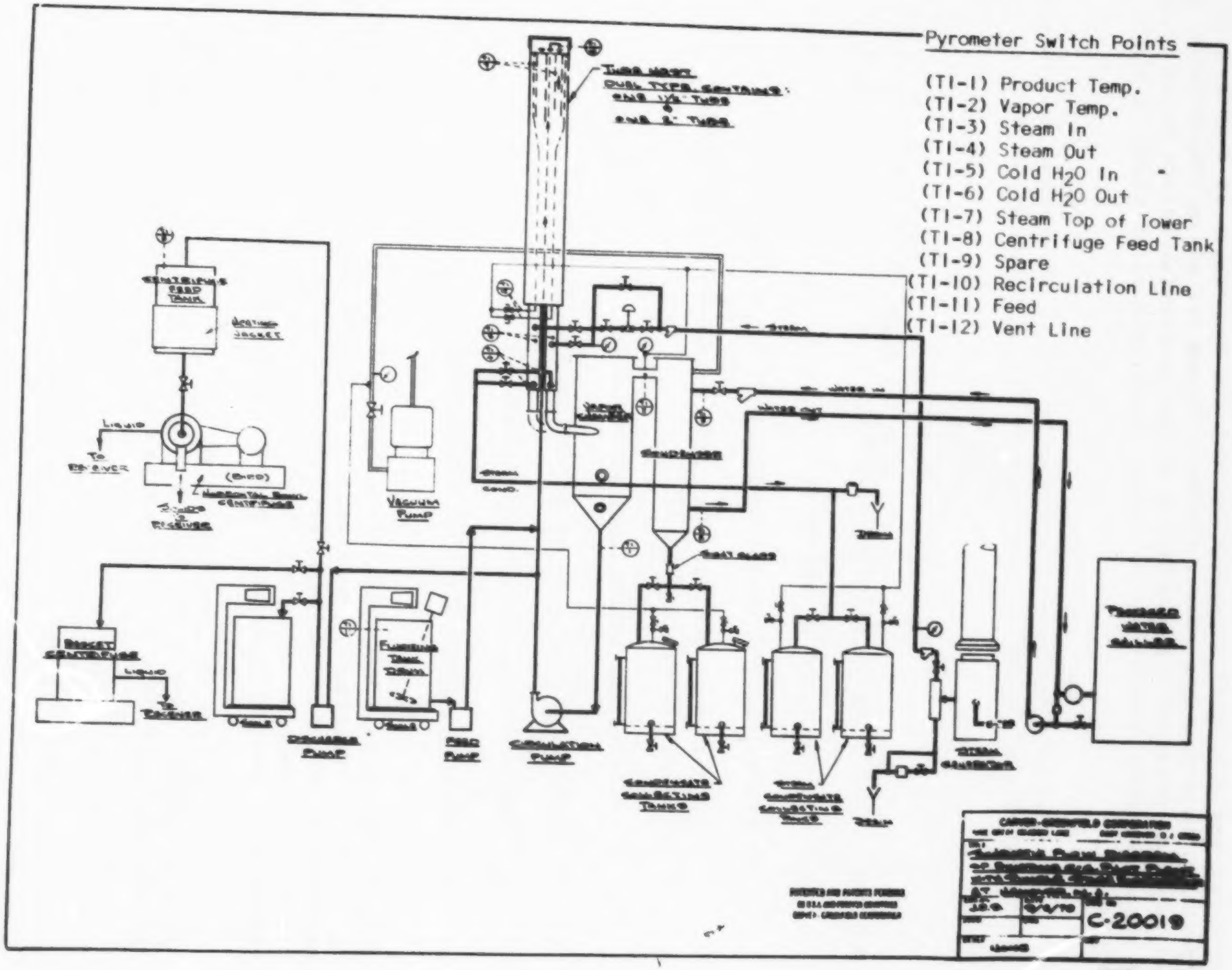
W_s = mass of steam condensate lbs/hr

H_s = heat of vaporization of steam Btu/lb

Thus U, the overall heat transfer coefficient (equal to the reciprocal of the overall heat transfer resistance) can be obtained by the experimental measurement of the steam condensate rate and temperature, and boiling sludge-oil mixture temperature during steady state operation. (The latent heat of vaporization and the evaporator heat transfer area are known fixed quantities.)

This type of test procedure, for determining U value, deoiling characteristics, distillate quality, etc. has been used routinely by Carver-Greenfield to get data for the design basis for commercial plants.

FIGURE 11


Pyrometer Switch Points
(TI-1) Product Temp.
(TI-2) Vapor Temp.
(TI-3) Steam In
(TI-4) Steam Out
(TI-5) Cold H_2O In
(TI-6) Cold H_2O Out
(TI-7) Steam Top of Tower
(TI-8) Centrifuge Feed Tank
(TI-9) Spare
(TI-10) Recirculation Line
(TI-11) Feed
(TI-12) Vent Line
Centrifuge Feed Tank
Heating Jacket
Liquid
To Receiver
Horizontal Bowl Centrifuge
Vacuum Pump
Basket Centrifuge
Liquid
To Receiver
Discharge Pump
Feed Pump
Circulation Pump
Steam
Water In
Water Out
Condenser
Drain
Condensate Collecting Tanks
Steam Generator
Drain
Carver-Greenfield Corporation
C-20019

6.2.1 High Viscosity Reduces Heat Transfer Coefficients in Initial Tests

Carver-Greenfield has carried out pilot plant testing of mixed primary and secondary sludges taken directly from two sewage plants: Hershey, Pa. and Bergen County, N.J.; in the test work on these plant samples the sludges were not preconcentrated further or otherwise treated, but mixed with the oil just before processing in the Carver-Greenfield evaporator. The Bergen County plant is the one which provided most of the sludge samples for the Esso program. These results provide a basis for comparison with the data for the Esso oil-sludge concentrates.

In the initial tests to simulate the first stage evaporation, the U values obtained with the viscous Esso oil-sludge concentrates were considerably lower than expected, in the range of 9-50 BTU/hr/ft^2/°F vs. ~120 for the directly processed sludges. The Carver-Greenfield test data are summarized in Table 27, with complete test data in Appendices C-1 and C-2.

The U values varied considerably, with no apparent effect of sludge type, sludge source, percent solids in the feed, or oil/sludge ratio. For the single test with #1 Varsol as the oil, the U values obtained were higher than for any of the tests with #4 heating oil by ~20%; further tests would be needed to confirm this difference, which could be an important factor in minimizing evaporator costs.

Because of the high viscosity, a simulation of three evaporation effects could not be made. Instead, the first stage concentrate was used as feed for the third stage (drying stage). U values for the Esso batches in the drying stage were lower than for the untreated sludge runs, but the differences were considerably less than found in the first stage (see Appendix C-2 for summary of drying stage data).

Carver-Greenfield's opinion was that the very high viscosities developed by the Esso samples during concentration were the major cause of the difference. This high viscosity drastically reduced the recirculation rate for the evaporator (estimated at 1/5 normal), due to the capacity limitations of the specific equipment in the Carver-Greenfield pilot plant. At these low recirculation rates the film resistance to heat transfer is drastically increased relative to normal flow. In addition, all of the transfer surface may not be utilized with high viscosity. The fluid being processed probably is not uniformity distributed over all the heat exchanger tubes in the desired thin film, but rather as thick films over some of the tubes only.

Carver-Greenfield was able to confirm the effect of flow rate in one test (which could not be duplicated) using chemical treatment to reduce viscosity. When recirculation rate was increased from an estimated 1 gpm to an estimated 3 gpm, the U value increased from ~10 to ~50. The U value projected for the normal pilot plant recirculation rate was estimated at 75.

TABLE 27

CARVER GREENFIELD HEAT TRANSFER TEST RESULTS

Run No.	Sludge Type and Source	Oil Used	Overall Heat Transfer Coefficient (U)	
			1st Stage	Drying Stage
1	Bergen County Activated	#4 Heating Oil	22-49	54-58
3	Bergen County Activated	#4 Heating Oil	23-32	33-45
4	Wards Island Activated	#4 Heating Oil	9-38	75-97
5	Wards Island Activated	#1 Varsol	41-60	50-110
6	Bergen County Prim. + Act.	#4 Heating Oil	19-82	
7	Trenton Trickle Filter	#4 Heating Oil	8-33	
Hershey, Pa Primary + Trickle Filter		Coray 37 (1)	93-122	93-132
Bergen County, N.J., Activated		#2 Heating Oil (1)	75-186	93-130

---

(1) Oil added just before evaporation; no prior processing.

The viscosity of the Esso concentrates is apparently much higher than the untreated sludges at the same solids concentrations. The explanation for this difference is not presently known, but is believed to be related to the solubilization/decomposition of the sludge solids prior to evaporation (see below). A series of tests were made in an effort to reduce viscosity by "demulsifying" the system, using pH adjustment and chemical treatment; none of the treatments tried was successful. An alternative approach, described below, was successfully developed by Carver-Greenfield.

Since the viscosity of the Varsol is considerably lower than #4 heating oil, it seemed reasonable to assume that the viscosity of an oil-sludge concentrate prepared with Varsol would be lower than with #4 heating oil. The viscosity data for comparable oil sludge concentrates using a Brookfield viscometer did not show this effect, however; since the concentrates are o/w emulsions, the viscosity of the continuous aqueous sludge phase appears controlling.

Oil in Concentrate	Average Viscosity (Brookfield) of Concentrate
#4 Heating Oil	380 cp
#1 Varsol	430 cp

Varsol, which has a relatively low boiling point (319-380°F), steam distills readily during the evaporation. The added turbulence due to the boiling oil may explain the higher U value compared to #4 heating oil; recent data from Carver-Greenfield with another type of sludge and a low boiling oil tends to confirm this hypothesis.

6.2.2 U Value Markedly Improved by Solids Recycle

Carver-Greenfield previously found that concentrate viscosity increased up to ~25% solids, then markedly decreased. This effect was successfully utilized to get around the viscosity-flow rate problem; centrifuged dry solids were added back to the feed sludge to produce 30% solids concentration, at which point the viscosity dropped appreciably. With this technique Carver-Greenfield reported close to normal recirculation rates, and a U value more than twice as high as previous results in the first test of the technique (run 7). A repeat test of the solids recycle technique, using the last batch of oil sludge concentrate from the Esso pilot plant program, was therefore made with the objective of confirming the higher U value for plant design (run 6); test results are shown in Table 28 below:

TABLE 28

SOLIDS RECYCLE IMPROVES U VALUE

	Overall U Value for 1st Stage	
Run No.	No Solids Recycle	Recycle to 30% Solids
6	19-52	42-66(1)
7	8-33	32-75

---

(1) Decomposed during long storage between tests.

Results of this final test (run 6) confirmed the effectiveness of the solids recycle, even though the improvement was not as great as the initial test. While the viscosity was reduced as expected, an improvement in U value of only ∼30% was obtained. Carver-Greenfield reported that the batch had "decomposed" during the long storage time (about 8 weeks) between the preparation of the oil-sludge concentrate and the heat transfer test; this decomposition could have changed the characteristics of the system sufficiently to effect the heat transfer; for example, poor dispersion of the sludge solids, or greater adherance of solids to the heat exchange surface would adversely effect the U value.

For the design of the evaporator system Carver-Greenfield used a U value of 60 BTU/hr/ft^2/°F; this was based on the last two test results and provides a reasonable basis for plant design, considering all factors. This U value of 60, which was attained only with solids recycle to reduce viscosity, is about 1/2 the value Carver-Greenfield obtained for sludges from the Hershey, Pa. and Bergen County, N.J. sewage plants. These sludges were not treated prior to the evaporation test, and therefore were not subject to thermal decomposition-solubilization of the solids prior to the heat transfer studies.

Solids recycle is considered completely practical for commercial operation by Carver-Greenfield, with an estimated small effect on overall economics. This technique of solids recycle was therefore used in their plant design studies based on the actual experimental data obtained.

6.3 Distillate TC Losses for Pilot Plant Batches

Distillate samples from all Carver-Greenfield runs were analyzed for nitrogen (as NH_4) and/or total carbon (TC), with available data summarized in Table 29. The Carver-Greenfield tests were carried out as a two stage evaporation process due to the operational problems in their pilot plant caused by high viscosity or solids recycle; about 2/3 of the initial water was evaporated in the first stage to simulate the first 2 stages of a 3 stage commercial unit, with the final 1/3 of the water removed in the drying stage.

TABLE 29

ANALYSIS OF DISTILLATES FROM CARVER-GREENFIELD HEAT TRANSFER TESTS

Esso Run	Distillate Stage (1)	pH	ppm as NH_4-N	ppm as Total C	TC/N	Odor
1	1st	8.9	160	265	1.65	Slight NH_3
	Drying	7.0	2400	6800	2.84	Petroleum, putrid
3	1st	9.1	330	480	1.45	Petroleum, putrid
	Drying	7.0	2000	5200	2.60	
4	1st	9.4	420	670	1.60	Petroleum, putrid
	Drying		3190	8950	2.81	
5	1st	9.2		460		Petroleum, putrid
	Drying	6.7				
6	1st	9.2		190		Petroleum, putrid
	Drying	5.5		13000		
7	1st	8.7		185		Petroleum, putrid
	Drying	N.A. →				

---

(1) Volume of 1st stage distillate stage $\sim$2x stage.

Both nitrogen and TC losses in the combined distillates were substantial, with the TC loss averaging 10% of the feed solids on a carbon basis; of this total, an average of 90% is contributed by the drying stage distillate, as shown below:

TABLE 30

TOTAL CARBON LOSSES IN CARVER-GREENFIELD EVAPORATION

Esso Run	#H_2O/#Solids in Feed	Wt. % Carbon in Feed Solids	Average PPM Total Carbon in Distillate(1)	Total TC Loss - %(2) in Total Distillate	Total TC Loss - %(2) in Drying Stage
1	13.8	34.8	2450	9.6	8.8
3	10.5	34.1	2100	6.5	5.4
4	13.7	38.9	3420	12.1	10.7
6	10.9	37.9	4560	13.0	12.6
			Average	10.3	9.3

---

(1) Data from Table 29, with ratio of 1st stage/drying stage volumes 2/1.

(2) Based on total carbon in sludge before oil concentration.

These losses in the drying stage distillate were unexpected, being ~4x greater than the values found for the distillates from the Hershey, Pa. sewage plant, where mixed primary + secondary sludge are dried in a three effect Carver-Greenfield evaporator system.

TABLE 31

TOTAL CARBON LOSSES IN DRYING STAGE DISTILLATE

Distillate Fraction	Average TC Content - PPM: 175°F Settling Esso Pilot Plant	Average TC Content - PPM: No Pretreatment Hershey Plant	% of Feed Solids in Distillate (1): Esso Pilot Plant	% of Feed Solids in Distillate (1): Hershey Plant
1st + 2nd Stage	410	520	1	1
Drying Stage	8500	2180	9	2

---

(1) Total Carbon basis.

While the distillates have not been analyzed to identify specific compounds, several general observations can be made:

- The high pH and ammonia odor of the first stage distillates indicates free NH_3 and or low molecular weight amines.
- The neutral-slightly acidic pH of the drying stage distillates, plus the high nitrogen content, suggests the presence of ammonium salts of volatile fatty acids.

- The decomposition of sludge solids and/or solubilized water soluble compounds (formed during the settling) appears very temperature sensitive. Only a minor portion of the total distillate loss (3-17%) occurred in the first stage evaporator at 130-160°F, with the major loss in the drying stage at 240-250°F.

A precise explanation for the high distillate losses relative to the Hershey plant data cannot be given at this time. Since the Hershey sludge was not treated prior to evaporation, a reasonable explanation, consistent with all the available data, is that the high temperature settling step used in the Esso process is responsible; either the water soluble compounds formed during settling are relatively low in molecular weight and therefore steam distill, or these compounds are unstable at the 250°F temperature in the drying stage and degrade further into compounds which do distill. The Hershey data demonstrate that high distillate losses are not inherent in the evaporation process per se. The effect of the time delay between the Esso concentration runs and the Carver-Greenfield heat transfer tests cannot be evaluated.

6.4 Composition of Product Streams from Evaporator

6.4.1 Analysis of Dry Solids

The dried sludge solids, with most of the oil removed in the centrifuge, were analyzed for residual oil and water; these solids were also deoiled by solvent extraction and the oil free residues analyzed for ash, carbon, hydrogen and nitrogen, and the heats of combustion were determined for use in heat balance calculations. Results are summarized below, with the detailed analytical data tabulated for the centrifuged and the deoiled solids presented in Appendix C-4:

- Residual oil in centrifuged solids: 40-47%, which is equivalent to 0.7-0.9#oil/# sludge solids. According to Carver-Greenfield these values are in line with their results on other sludges processed.

- Residual water in centrifuged solids: 0.9-2.8%, with an average of 1.6%. Again, Carver-Greenfield reports that those values are consistent with results from other sludges.

- Ash content of deoiled centrifuged solids: 40-47% vs. 30-37% for the deoiled, initial feed solids (see Table 2).

- C,H,N content of deoiled centrifuged solids: 25-38%, 3.8-5.5%, 4.0-4.9%, respectively for the three elements.

The significantly higher ash values for the final solids, and the correspondingly lower C,H,N values, reflect the TC & TOC losses in the raffinates and distillates during processing.

- Heat of combustion (heating value): net heating values of 4,570 BTU/# and 4,850 BTU/# were obtained for the oil free centrifuged solids from Bergen County and Wards Island activated sludges. The average of 4,680 BTU/# was used for the heat balance calculations for the Carver-Greenfield process; this value is ∿16% lower than calculated for the feed sludge, and reflects the reduction of combustible (volatile) components due to TC losses.

The centrifuged dry solids from the Esso concentrates were odorless, dark brown, crumbly solids; these solids had an appearance and physical "feel" very similar to Michigan peat moss.

6.4.2 Analyses of Recycle Oil

Oil recovered from the centrifuging step was evaluated for water, sludge solids not removed in the centrifuge, and viscosity. Test data are tabulated in Appendix C-5 and can be summarized as follows:

- Water content: 0.1%
- % non fat sludge solids: 1.9-2.7%
- Viscosity: 116 Saybolt seconds (100°F) vs. 73 for fresh oil.

The residual solids level in the centrifuge oil is normal for the Carver-Greenfield process. According to Carver-Greenfield the solids level in the recycle oil will not build up beyond this value, since an equilibrium condition is rapidly established.

The viscosity increase in the recycle oil is probably attributable to the solids content. No pumping or fluid flow problems have developed in any of the commercial Carver-Greenfield plants; based on this experience, no problems would be expected for a plant operating the Esso-Carver Greenfield process, either.

6.5 Reduced Concentration Temperature for Esso Process is Indicated

Comparing the results of the heated Esso sludges and the untreated sludges, a major reduction in the severity of heat treatment, as measured by TC loss in the raffinate, should significantly improve the heat transfer properties of the Esso oil sludge concentrates. One way to accomplish this would be to eliminate the 175°F settling stage and operate at 105°F. The reduced settling temperature would cut TC loss in the raffinate from ∿15% to ∿7% and should similarly reduce the distillate TC loss from ∿10% to ∿3%.

The net expected effect of the lower temperature operation is an increase in the U value of up to 100%, to a value closely approaching that for sludges not processed thru the Esso oil concentration step.

Operation at a lower settling temperature would have other effects on the overall process, the most important ones being reduced heating requirements for the oil sludge concentration step and a decrease in the final solids concentration. Detailed evaluation of this process alternative has been made as part of the Phase 4 program.

7. PHASE 4: PROCESS TRADE OFF STUDIES AND COST ANALYSIS

7.1 Process Flow Plan for Commercial Plant

In order to simplify the task of cost estimation and process optimization, the overall process was divided into sections or modules, for separate detailed analysis. The primary modules considered were:

- Waste sludge thickening.
- Oil-sludge mixing and concentration (settling).
- Multiple effect evaporation of oil sludge concentration.
- Incineration of the dry solids.

The first two represent the "Esso Process", and the last two the "Carver-Greenfield Process". The detailed cost estimation, equipment sizing, etc. for the Carver Greenfield steps were made by the Carver Greenfield Corp.; modifications were made as required by Esso for integration with the Esso modules and to evaluate several of the process alternatives considered.

A further simplification in the overall process and cost analysis was to design the Esso process modules for secondary sludge alone. The design basis for the Carver Greenfield process has been set up with wide enough limits to permit inclusion of primary solids at a later date by merely adjusting the total water and solids load. Since dewatering of secondary sludges to a concentration suitable for feeding to the Carver Greenfield process is much more difficult than primary sludges, the final integrated system will require processing of the secondary sludge alone thru the Esso concentration process, then mixing with conventionally thickened primary sludge for evaporation and incineration.

A listing of the actual individual steps in the overall process and their interrelation will provide a useful introduction for the design basis and the specific assumptions used in the process studies and cost estimates. The process flow plan used for the basic design is shown in Figure 12 and consists of the following steps:

- Prethickening of waste secondary sludge from 0.5% to 1.5% in a conventional thickener.
- Preheating the thickened sludge to 105°F by direct contact in the barometric condenser of the 3rd effect of the Carver-Greenfield evaporator system.
- Mixing the preheated sludge with hot recycle oil from the Carver-Greenfield process in a high shear, in-line mixer.
- Settling of the oil-sludge mixture in three separate stages.

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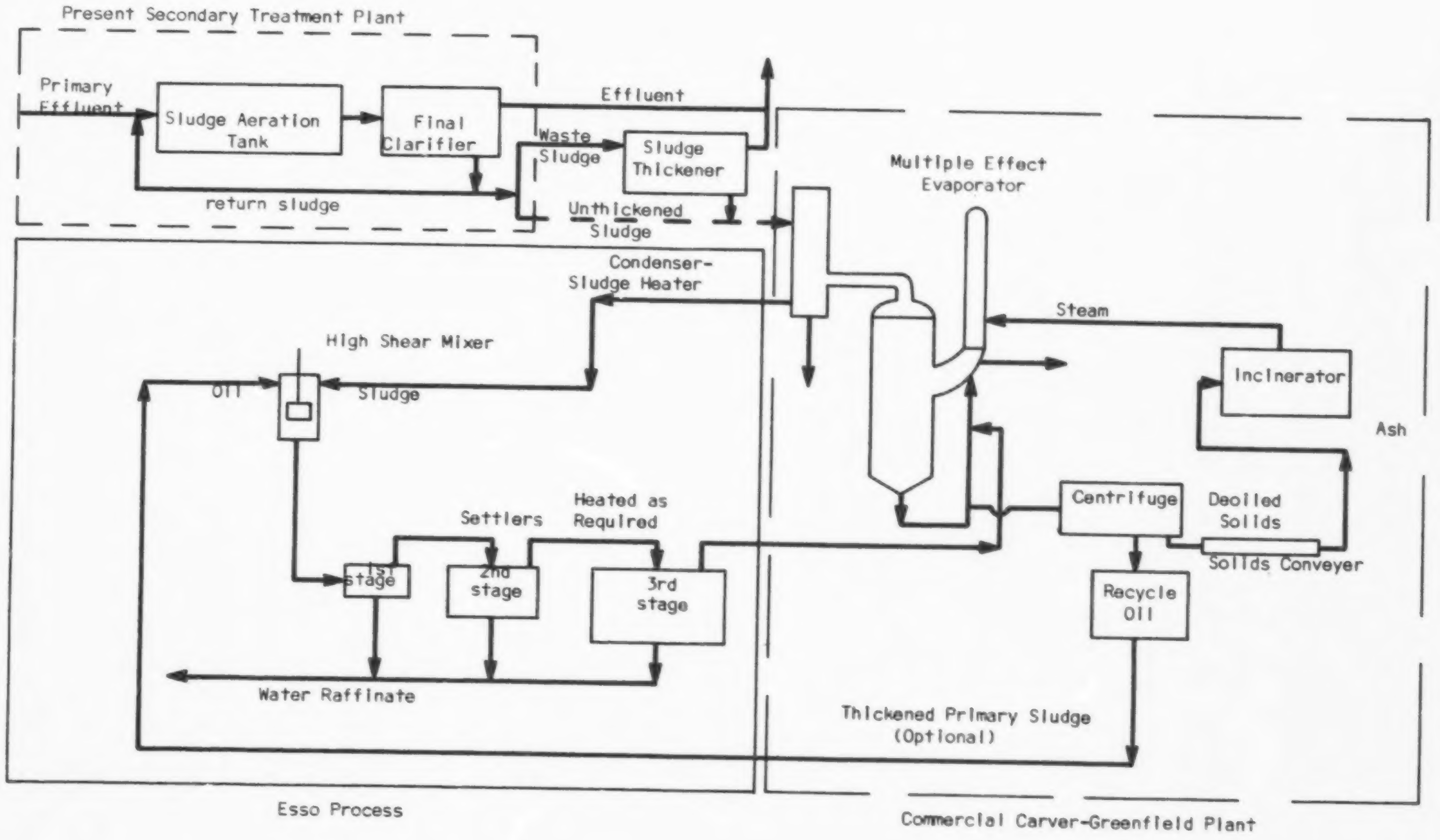
FIGURE 12

SCHEMATIC FLOW PLAN OF ESSO-CARVER GREENFIELD PROCESS


Present Secondary Treatment Plant
Primary Effluent
Sludge Aeration Tank
Final Clarifier
Effluent
Waste Sludge
Sludge Thickener
return sludge
Unthickened Sludge
Multiple Effect Evaporator
Condenser-Sludge Heater
Steam
High Shear Mixer
Oil
Sludge
Incinerator
Ash
Centrifuge
Deoiled Solids
Solids Conveyer
Settlers
Heated as Required
1st stage
2nd stage
3rd stage
Recycle Oil
Water Raffinate
Thickened Primary Sludge (Optional)
Esso Process
Commercial Carver-Greenfield Plant

  - 1st stage, which is characterized by rapid separation of phases, in a horizontal drum with 1 hr. residence time.
  - 2nd stage in a covered sludge thickener, but without the "picket-fence" agitator, in which the oil-sludge concentrate is taken off as a "float" phase.
  - The effluent from the 2nd stage is then heated to 175°F for the 3rd and final stage; this 3rd stage settler is the same type as the second stage unit.
  - Water raffinates from the 3 settling stages, containing solubilized feed solids, are recycled to the inlet of the secondary treatment plant.

- The final oil-sludge concentrate from the final settling stage is then fed to the Carver-Greenfield multiple effect evaporator system (3 or 4 effects, reverse flow). Primary sludge is mixed with the concentrate before the evaporation step, where mixed sludge is processed to dryness.
- After the final drying stage, the dry solids are separated from the oil in a Bird solid bowl centrifuge; the oil is recycled to the Esso concentration step and the solids deoiled by solvent stripping to the extent required for heat balance.
- The deoiled solids are then burned in an incinerator with the heat energy used to generate steam for the multiple effect evaporator system.
- As the final step in the process, the ash from the incinerator is removed for disposal; the quantity of ash will be ~30-35% by weight of the input secondary sludge solids to the concentration process. The actual volume of ash is <0.1% of the waste sludge volume, with 0.5% solids content assumed.

7.2 Basis for Process Design and Analysis

Brief descriptions of the design basis used for sizing the major units in each step of the Esso concentration process are given below. The process design, heat balance and cost data for the complete, installed Carver-Greenfield process were provided by Carver-Greenfield for 9 different cases (see Appendices D-6, 7 and 8). These cases covered the range of plant sizes and concentrate feed solids contents selected for the process analysis, as well as evaluating the effect of increasing U value on plant cost; the cost data for all other cases required to complete the study were computed using the basic design and cost data provided by Carver Greenfield.

Equipment cost data were obtained from the open literature, updated using appropriate inflation indices. Factors for estimating capital costs (amortization and interest), maintanence, insurance, total labor rate (direct operating, overhead, supervision, etc.), engineering fee and contingency factor were obtained from the literature and discussion with knowledgeable persons in the field. The various factors and cost basis used in the estimating are summarized in Table 32, as of March, 1972.

Equipment sizing was based on 100% of design capacity in all cases, and on 365 days/year operation.

TABLE 32

COST FACTORS AND INDICES USED IN ESTIMATES

- Inflation Factors for Equipment: US Dept. of Commerce Composite Index, = Sewage Treatment Plant Construction Cost Index (41, 44).
- Amortization Rate: 25 year straight line.
- Interest Rate: 5% (Feb. 1972 rate on AAA Bonds).
- Engineering Cost: See Appendix D-1 for variable rates (37).
- Contingency Factor: 10% of installed equipment cost.
- Maintanence labor + materials: 5% of total erected + installed (TEI)* cost (34, 35, 38).
- Insurance: 1% of total installed cost (34, 35).
- Labor Rate: Direct Operating: \$3.90/hr (43)
  Indirect Operating: \$0.60/hr (37)
  Plant Overhead: 30% of Direct + Indirect labor = \$1.35/hr (37).
- Electricity: \$0.01/KWH (37, 39).
- Fuel (#4 Heating Oil): \$0.016/lb. (from Humble Oil & Refining Co.).

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* Installed equipment + engineering cost + contingency = TEI cost.

7.2.1 Selection of Plant Sizes

The range of plant sizes chosen for any process and cost analysis must be somewhat arbitrary, but also must be broad enough to cover the likely range of both interest and practicality. For this study, plants with raw influent flows of 12.5-250 MGD were selected; these plants are the size estimated to serve populations of about 100,000 to 2,000,000 people.

The waste secondary sludge was fixed at 1.8% of the plant influent volume, with a suspended solids content of 0.50%; these values correspond to secondary sludge solid feed rates of 4.72-94.5 tons/day.

The population equivalents to influent flow and the waste secondary sludge/plant influent ratio were based on literature references (4, 21) and were intended primarily as guides to plant size. All detailed cost calculations and equipment sizing was made on the basis of sludge solids feed rates, however.

For calculating plant sizes for mixed primary-secondary sludges, a 50-50 weight ratio was selected (48).

7.2.2 Prethickening of Waste Secondary Sludge

A total of 14 batches of secondary sludge have been processed from 3 different plants; 9 from Bergen County, 4 from Wards Island and one Trenton. Settling characteristics of all batches were measured by the standard 1 liter batch test. The settling rates varied appreciably for different batches from the same plant, as show in Table 33, listing the data for the slowest and fatest batches from Bergen County and Wards Island. The Trenton, N.J. trickling filter sludge fell within these ranges.

Design of the thickener, in terms of surface area required, was calculated by the Kynch method, using the procedure of Talmadge and Fitch (27). The actual initial suspended solids contents of the sludge batches as obtained from the plants was 0.5-0.6%; in all cases the thickener design was based on an underflow concentration of 1.5%, which was the maximum attainable for the slowest settling batches.

7.2.3 Heating of Thickened Sludge Prior to Oil Contacting

Heating of the sludge was carried out using the sludge as the condensing (cold) fluid in the barometric leg of the 3rd effect evaporator.

Including the heat input from the hot recycle oil, the total heat available for preheating the sludge is adequate even for winter conditions (40°F minimum sludge temperature); preheating of the oil sludge mixture to 115°F at the inlet of the 1st concentration (settling) stage was assumed. This heat balance was one of the considerations used for selection of the prethickening step to 1.5% solids.

TABLE 33

SECONDARY SLUDGE SETTLING CURVES

Ambient Temperature (20-25°C)
Initial Suspended Solids = 0.5 - 0.6%

Settling Time Hrs	Relative Sludge Volume (V/Vo) (1)	
	Bergen County (2)	Wards Island (3)
0	1.0	1.0
1	.27-.77	.55-.88
2	.215-.56	.36-.75
3	.195-.38	.32-.65
4	.188-.31	.30-.52
6	.173-.27	.292-.48
8	.168-.258	.289-.465
10	.166-.248	.285-.455
12	.164-.243	.283-.450
24	.155-.240	.275-.445

(1) Range of data for individual batches tested.

(2) 9 batches.

(3) 4 batches.

7.2.4 Oil-Sludge Mixing

Use of an in-line mixer, which is basically a turbine agitator with close clearance to the mixing chamber wall to provide high shear rates, was assumed. The costs for in-line mixers are comparable to those for centrifugal pumps actually used in the experimental program, so choice of the specific mixing system will have no effect on the cost estimate.

7.2.5 First Stage Settler

This was designed on the basis of a conventional, horizontal drum type oil-water separator, with 4 horizontal baffles to increase effective settling area; the rapid initial rate of settling permits use of this low cost unit for the first stage only.

7.2.6 Second and Third Stage Settlers

These were designed as "inverse thickeners", with the conventional rake agitator replaced by a top skimmer, similar to that used in an air flotation unit. The actual sizing was based on use of the Kynch method, since the settling curves for oil-sludge are very similar to waste sludge and the controlling mechanism was also judged very similar; literature references support use of this approach (49).

The oil-sludge concentrate from the second stage settler must be preheated to 175°F before the final settling stage. In the process design this preheating was based on heat exchange in a conventional shell and tube exchanger, using 100# steam as the heating fluid, with the steam generated in a self contained, "package" boiler unit.

7.2.7 Process Oil and Oil-Sludge Concentrate Storage

An oil inventory equivalent to 36 hours operation of the Esso concentration process was assumed, with the cost capitalized. Oil storage capacity equivalent to the oil inventory was included in the overall design.

To provide surge capacity for both the Esso concentration process and the Carver-Greenfield process, a storage tank for oil sludge concentrate, with 24 hour capacity, was provided for in the design.

7.2.8 Carver Greenfield Process

The design and plant cost estimation for the Carver-Greenfield process, including both the evaporation and incineration steps, was made by Carver Greenfield. Their cost estimates are for installed plants, with all equipment, installation, engineering and contingency included. The incinerator cost, which represents a large component of the total investment was based on the average of quotes from two different suppliers of commercial units, and also includes engineering plus contingency.

Costs were calculated on two bases: 1) using the experimentally determined U values found with the Esso pilot plant batches concentrated at 175°F, 2) using a U value 2x larger, equivalent to values obtained for sewage sludges without prior treatment, and which we assumed would also be obtained with Esso sludges concentrated at 105°F. In the first case 3 effect evaporation was assumed; in the second high U case, costs were calculated for 3 and also for 4 effects for those situations where 4 effects are practical - namely $\leq$ 6% solids in the feed and plant sizes sizes > ~14 tons/day.

7.2.9 Esso Oil Concentration Process

Sizing of settlers, pumps, heat requirements, etc. was based on the experimentally determined concentration factors found in the batch laboratory process studies and the pilot program. Process calculations were made for the maximum, minimum and average concentration factors found experimentally for the particular sludge feed suspended solids content and settling temperature-time cycle selected as the base conditions. These conditions were selected as follows:

- % suspended solids in feed: 1.5%
- laboratory settling conditions: 5 hrs. at 105°F
  15 hrs. at 175°F

For these settling conditions, the values for concentration factors taken from the experimentally derived curve were 8, 6, 4 (see Figure 3). This curve summarizes all of the experimental data, including pilot plant results, of concentration factor vs. % feed solids for 20 hours at 175°F; the most recent test data showed that essentially the same results were obtained for the 5 hr. at 105°F - 15 hrs. at 175°F cycle. To convert the initial feed solids concentration to final solids in the concentrate, the concentration factors must be corrected for the following:

- Weight loss due to Total Carbon (TC) losses in the raffinates = 15% of initial weight. This value is based on the average TC loss found experimentally for the settling conditions above, and assuming that the total loss of solids weight equals the loss of TC.
- Oil solubles in the feed sludge solids = 10% of initial weight. This value represents the average oil soluble content of 4 sludge batches; the correction must be made since any oil soluble fraction present in the sludge solids will remain in the oil even after evaporation of the water in the Carver Greenfield process.

Using the concentration factors specified above and the two correction factors for TC loss and oil solubles in the sludge, the solids contents after the Esso concentration are 4.5, 6.0 and 9.0% respectively.

7.3 Procedure Used for Cost Estimates

As noted above, the Esso process involves two primary and essentially independent steps - prethickening of the sludge and oil concentration. The widely varying thickening rates of the waste sludge mean a wide range of thickener surface area requirements (loadings of 2.5 - 10.4 lbs/day/ft^2), and therefore thickener costs.

Similarly, for a fixed feed solids content to the oil concentration step, (1.5%) a wide range of final solids concentrations will be attained. Based on the actual experimental data obtained in the program to-date, this concentrate solids content, which is the feed to the Carver Greenfield process, can vary from 4.5-9.0% (equivalent to concentration factors of 4-8).

A relationship between settling rate of the waste sludge and the concentration factor attained in the oil-extraction process has not been found for batches from a particular plant. However, comparing the two plants tested in the program, Bergen County and Wards Island, the available data clearly show that:

- Settling rates of waste sludges (for thickening) from Bergen County are higher than from Wards Island.
- Concentration factors of batches from Bergen County are also higher than from Wards Island.

For purposes of broad cost analysis, then, the fastest settling sludge batches can be associated with those batches giving the highest concentration factors and vica versa. The cost analysis for the Esso process was therefore set up for three sets of process combinations to permit coverage of the full range of practical possibilities:

- Fastest thickening + highest concentration factor (9.0% solids)
- Medium thickening + medium concentration factor (6.0% solids)
- Slowest thickening + lowest concentration factor (4.5% solids)

The final cost composite for the combined Esso-Carver Greenfield process is straight forward, since the C-G process is calculated separately on the basisi of 4.5, 6.0 and 9.0% solids in the feed to the evaporators.

Two types of cost estimates were made: a) processing only secondary sludge, such as might be the situation with a contact stabilization plant, or if some alternate process were available for primary sludge, b) processing only secondary sludge thru the Esso process, and combining the concentrate with primary sludge for the Carver Greenfield evaporation and incineration steps.

For secondary sludge alone, the size of both processes is fixed only by the initial feed solids load; combined total cost data for the Esso-Carver Greenfield processes is therefore appropriate. For mixed primary + secondary, the solids load for the Carver Greenfield process will be greater than for the Esso process, with the factor dependent upon the weight ratio of primary/secondary. Separate cost curves for the Esso process and for the Carver Greenfield process were set up to permit selection of the components for the assumed 50/50 weight ratio and also for differing ratios as required.

All cost estimates were prepared on the basis of two sets of process responses:

- Present Data Basis - using the experimental data actually obtained for the concentration and evaporation steps in Phases 1-3.
- Projected Data Basis - using the results expected from modified operating conditions.

7.4 Cost Estimates - Present Data Basis

7.4.1 Esso Oil Concentration Process Component

Capital costs for the Esso process components covering the range of plant sizes, waste sludge settling rates, and concentration factors discussed above are summarized in Appendix D-2. Costs for the individual steps of sludge thickening, oil-sludge mixing, oil-sludge heating before 3rd stage settling, oil-sludge settling, concentrate surge tank, process oil storage, and process pumps are included.

Total investment for the Esso process, and the detailed summary of the individual components of the treatment costs, are given in Table D-3 and summarized below:

TABLE 34

COSTS FOR ESSO PROCESS COMPONENT (1972)

Tons Sludge/Day	Total Investment, $MM		Total Capital and Operating Costs $/Ton Sludge	
	Low	High	Low	High
4.72	.32	.38	48	54
14.12	.55	.69	27	32
94.5	2.0	2.8	18	22

Low = fastest thickening, highest concentration.
High = slowest thickening, lowest concentration.

The major fraction of the total investment, and therefore of the operating cost as well, is tied up in sludge thickeners and oil-sludge settlers:

TABLE 35

INVESTMENT BREAKDOWN FOR ESSO PROCESS COMPONENT

	% of Total Investment		
Tons Sludge/Day	Sludge Thickening Average	Oil Sludge Concentration Average	Combined
4.72	24	39	63
14.16	24	41	65
94.5	26	44	70

Low = fastest thickening
High = slowest thickening

Since thickeners and settlers represent such a large component of investment and total operating cost, there is considerable incentive to reduce the size (surface area requirements) of these units. As will be discussed later in greater detail, there are several possible approaches to significantly reducing equipment size and correspondingly improving process economics.

7.4.2 Carver-Greenfield Process Component

The consolidated total investment and total operating costs for the Carver Greenfield process are tabulated in Appendix D-4. Three different sets of process conditions and/or equipment design bases were evaluated and costs calculated:

1. Experimentally determined overall heat transfer coefficient (U value = 60 BTU/hr/ft^2/°F), experimentally determined average Total Carbon (TC) losses in raffinate and distillate, (15% TC in raffinate, 10% in distillate), 3 effect evaporator system. This set of process conditions will be referred to as the "Present Data Basis".

2. U value of 120, reduced Total Carbon loss (7% in raffinate, 3% in distillate), 3 effect evaporator system. The U value and TC losses are the projected values considered attainable with low temperature settling in the Esso concentration process. These assumed conditions represent the "Projected Data Basis".

3. Same as 2, but with 4 effect evaporator system in place of 3 effect. According to Carver-Greenfield, use of a 4 effect evaporator is limited practically to ∿6% maximum feed solids content and ∿14 tons/day minimum sludge solids rate.

Total capital and total operating costs for the Carver Greenfield process, using the present data (experimental) basis, show a large cost reduction with increasing plant size and increasing feed solids content.

TABLE 36

COSTS FOR CARVER GREENFIELD PROCESS COMPONENT

Plant Size Tons Sludge/Day	% Solids in Feed	Total Investment $MM	Total Capital and Operating Cost $/Ton Sludge
4.72	4.5	.62	84
	9.0	.56	73
14.16	4.5	.88	40
	9.0	.64	26
94.5	4.5	3.8	27
	9.0	2.4	12

The unusually large effect of increasing solids content for the 94.5 tons/day plant size is mainly due to reduction in the number of individual evaporator trains required (from 4 to 2), as well as a reduction in manpower (from 3 to 2 men/shift) and the size of the incinerator-boiler: the cost of the incinerator-boiler is largely dependent upon the steam load.

7.4.3 Combined Esso Carver Greenfield Process Secondary Sludge

The individual investment and operating costs for the Esso and Carver-Greenfield process components are shown in Figures 13 and 14 based on the present experimental data. These cost estimates are based on continuous (3 shift) operation for all size plants. For the limiting combination of process conditions for the Esso concentration process, the total investment, total operating costs, and operating cost breakdown are summarized below:

FIGURE 13

TOTAL INVESTMENT (TIE) FOR ESSO AND CARVER GREENFIELD PROCESS COMPONENTS - PRESENT DATA BASIS

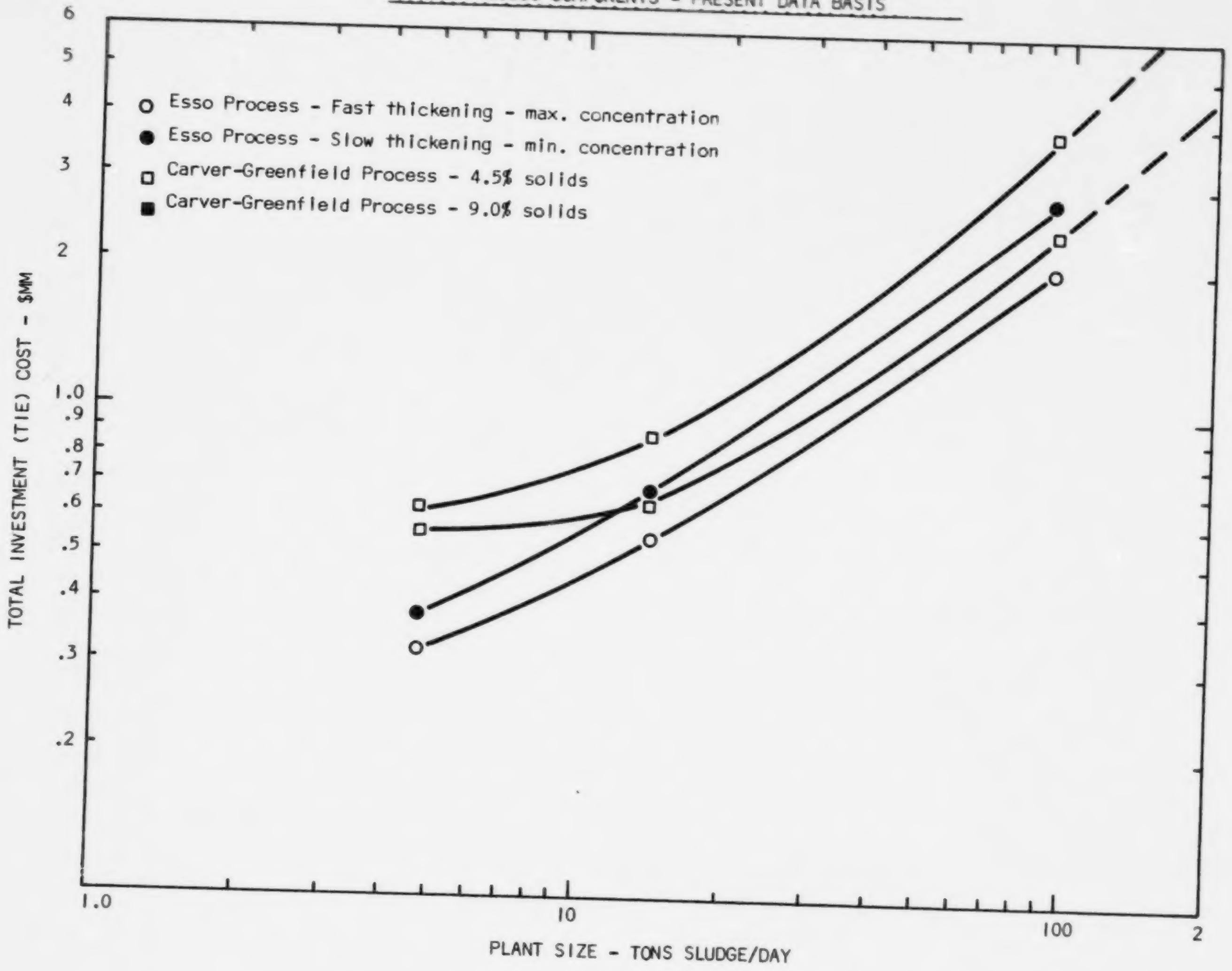

FIGURE 14

TREATMENT COSTS OF ESSO AND CARVER GREENFIELD
PROCESS COMPONENTS-PRESENT DATA BASIS

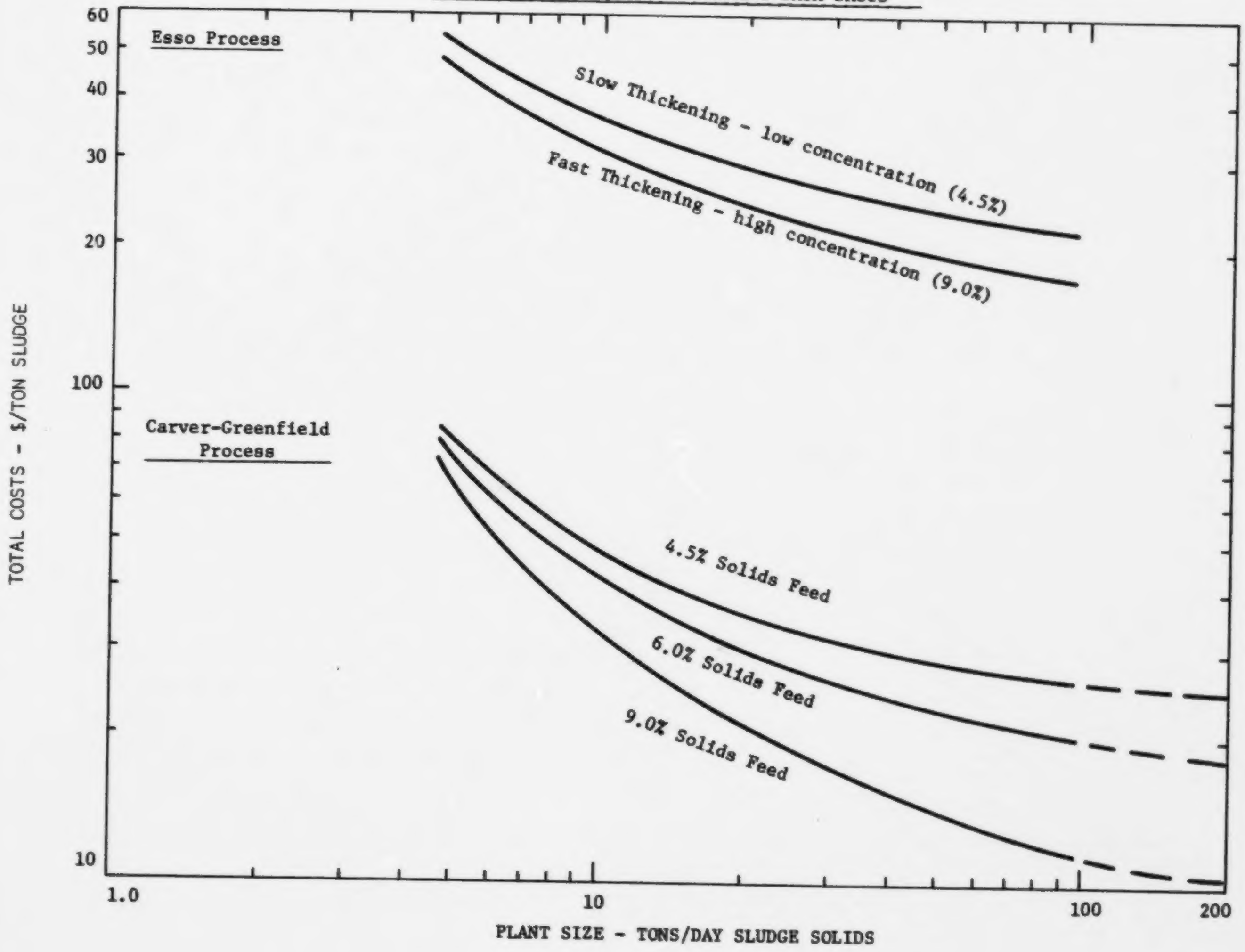

TABLE 37

COSTS FOR COMBINED ESSO-CARVER GREENFIELD PROCESS; SECONDARY SLUDGE ONLY

Plant Size Tons/Day	Process Conditions	Investment $MM	Total Cost, $/Ton	Total Cost Breakdown - % Based on			
				Investment	Labor	Power +Fuel	TC Loss
4.72	Max.	.88	121	55	37	4.1	3.6
	Min.	1.0	138	55	32	9.8	3.2
14.16	Max.	1.2	54	56	27	8.3	8
	Min.	1.6	73	55	21	18	6
94.5	Max.	4.4	29	57	15	13	15
	Min.	6.6	49	52	12	27	9

Max = fastest sludge thickening, highest concentration factor, 9% solids to evaporator

Min = slowest sludge thickening, lowest concentration factor, 4.5% solids to evaporator.

The effect of size on investment deviates from the normal for the Carver Greenfield process at the small-size end of the curve; the difference between the 4.7 T/D and 14.2 T/D plants is small, particularly for the 9% solids case. According to Carver Greenfield, the costs of the instrumentation, controls, auxiliaries and engineering remain essentially the same for the two size plants; as well as the actual cost differential for the incinerator-boiler.

Investment based costs are >50% of the total for all size plants, with labor based costs the next most important category. The inverse relationship between power + fuel, and TC loss vs. plant size reflects the fact that these items are only slightly dependent upon size.

7.4.4 Costs to Recycle Total-Carbon Losses

Total Carbon losses for the present data basis were assumed to be 25% of the feed sludge, for both the oil concentration and evaporation steps. The process was debited with a recycle "cost" of $4.37/ton, calculated from the BOD equivalent of the TC recycled and the influent charge for BOD to the Chicago municipal sewage system (47); this is equivalent to $1.75/10% TC loss. Similar debiting of the operating cost for recycle loads associated with any process is required for valid comparisons of different competitive processes.

7.4.5 One Shift vs. Three Shift Operation

For the smallest size plants the labor costs component is relatively the highest. Operation of the plant on a one shift basis can under certain conditions, result in a significant cost reduction.

For one shift operation the plant must be sized 3x larger than for continuous operation; reduced labor costs must then be balanced against increased capital based costs.

For the Carver-Greenfield plant, processing 4.7 tons/day, there is a considerable cost advantage for one shift vs. 3 shift operation at the highest level of solids in the concentrate; there was no advantage for one shift operation at the lowest solids level, however. For the Esso process component, one shift operation was more costly than three shift:

TABLE 38

COST COMPARISON OF ONE VS THREE SHIFT OPERATION

Plant Thruput: 4.7 Tons/Day Sludge Solids

% Solids in Concentrate	Number Shifts	Total Costs, $/Ton Esso Process Component	Total Costs, $/Ton Carver Greenfield Process Component
4.5	1	63.5	84.8
	3	53.9	84.2
9.0	1	52.1	58.5
	3	48.1	73.1

The difference in response for the Esso and Carver Greenfield components is due to the differences in the equipment vs size cost relationship (see Figure 13). For the smallest size Carver Greenfield plant there appears to be a definite cost advantage for one shift operation except at the lowest solids content (most unfavorable process response).

One shift operation of the evaporation-incineration process is actually practiced at the Hershey, Pa. sewage treatment plant; no operating difficulties associated with one shift operation have been reported.

7.4.6 Combined Esso-Carver Greenfield Process

The majority of treatment plants require disposal of primary as well as secondary sludges. As noted above the design basis for these plants was set up on the assumption of processing only the secondary sludge thru the Esso concentration process, then adding the primary sludge for the Carver Greenfield process. In this way the capital investment and operating costs required for the Esso process is minimized and TC loss from processing of primary sludge avoided.

The actual plant size required, as well as the final investment and operating cost, will depend on two factors for any given plant: ratio of primary/secondary sludge solids and the concentration of the primary sludge; this latter factor in turn will depend upon the operation of the specific plant (type solids, operation of primary sedimentation unit) and whether or not a sludge thickener is available.

The EPA recommended a value of 50/50 for the primary/secondary sludge ratio. Based on a literature survey + personal references, the following values for primary sludge solids contents were assumed:

- no thickener: 4-8%
- thickener : 8-10%

Combining these solids contents with the range of solids from the Esso process for secondary sludge, (4.5-9.0%) the following design basis would be obtained, assuming 15% oil solubles in the primary sludge:

% Solids in Primary Sludge	Maximum Range of % Solids in Combined Feed to Carver-Greenfield
4-8	4.0-7.9
8-10	5.7-8.8

Using the 50/50 sludge ratio, total costs were calculated for a solids range from the Esso process (concentrated secondary sludge) of 4.5-9.0%, and a combined feed to the Carver Greenfield process of 4.0-8.8%. Costs for the Carver Greenfield process component were taken from the curves in Figure 15, and for the Esso component from Figure 14. Since the final mixed sludge contains only 50% activated, the cost for the Esso process component was multiplied by 0.5 in calculating its contribution to the total process cost.

For a new plant, or one without a thickener, the cost of primary sludge thickening must be added to the costs of the Esso and the Carver-Greenfield components to get the total process cost. The following costs for thickening primary to 8% (37, 48) were used for this purpose:

FIGURE 15

EFFECT OF % SOLIDS IN CONCENTRATE ON COST OF CARVER GREENFIELD PROCESS (3 EFFECT EVAPORATOR)

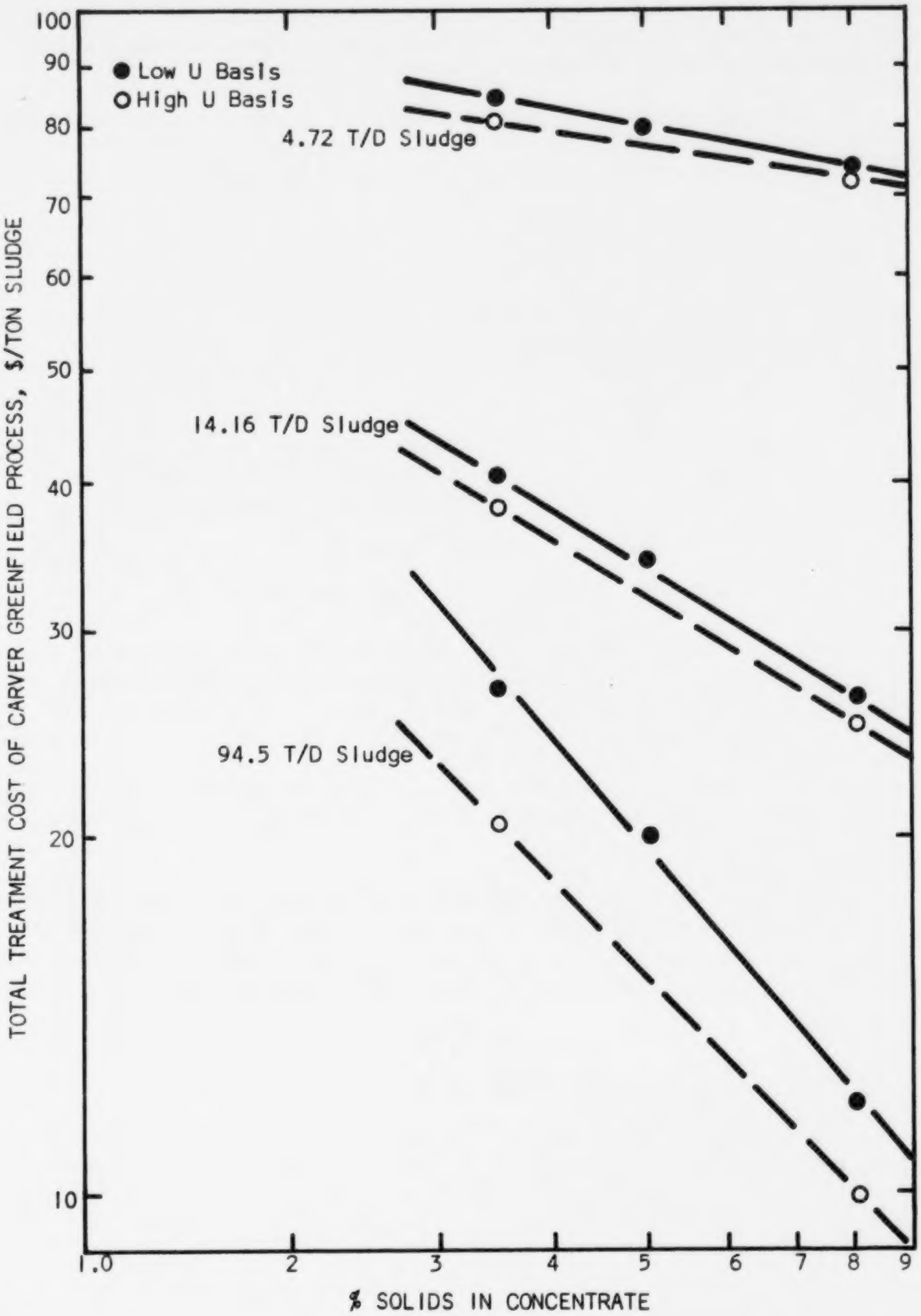

TABLE 39

COSTS FOR THICKENING PRIMARY SLUDGE

Sludge Load Tons/Day (1)	Thickening to 8% Cost - $/Ton (2)
4.7	3.60
14.2	1.90
47.2	1.10
94.5	.80

(1) Primary
(2) Primary + secondary (total sludge basis).

Total costs for the combined Esso-Carver Greenfield process, including the cost of thickening primary sludge, are summarized in Table 40 below:

TABLE 40

TOTAL COSTS FOR 50/50 PRIMARY + SECONDARY SLUDGE

Present Data Basis

Plant Size Sludge Load Tons/Day	Total Cost Range, $/Ton Esso	Carver Greenfield	Total Process
9.4	24-27	34-53	62-80
28.3	14-16	18-36	34-52
94.5	9.6-12	12.5-30	23-42
189 •	8.8-11	10.7-28	21-39

The values for the Carver Greenfield component of the largest size plant were obtained by extrapolation of the curves, and many therefore contain an added "uncertainty factor" estimated at ± $1/ton.

7.5 Cost Estimates - Projected Data Basis

Review of the cost components of the overall process, based on the present experimental and design basis point up several areas where large cost savings could be achieved:

- Reduction in Total Carbon losses.
- Increase in overall heat transfer coefficient (U).

- Reduction in fuel costs required to heat oil-sludge concentrate to final stage temperature of 175°F.

- Reduction in size (surface area) of oil sludge settlers for concentration step.

The first two items, TC loss and U value, are believed to be closely connected; TC losses are an indication of the decomposition/solubilization of sludge solids, believed to be responsible for the the low U values compared to sludges not heat treated prior to evaporation.

7.5.1 Lower Oil Sludge Concentration Temperature Expected to Reduce Costs Appreciably

The primary objectives for the use of a lower temperature for the oil sludge concentration step are a) reducing the TC losses in the concentration and evaporation steps and b) increasing the U value for evaporation. Use of a constant settling temperature of 105°F was considered the most reasonable choice on the basis of balancing both TC loss and the concentration factor: while total TC losses will be reduced by an estimated 60% compared to the base case of 175°F, concentration factor will be reduced by 10-25% at the same time, as summarized below in Table 41:

TABLE 41

COMPARISON OF PROCESS RESULTS FOR
105°F AND 175°F OIL SLUDGE SETTLING

Settling Conditions	Concentration Factors		TC Losses		Final Solids Content - %	
	Max	Min	Raffinate	Distillate	Max	Min
5 hrs 105°F, 15 hrs 175°F	8	4	15	10	9.0	4.5
20 hrs 105°F	5.4	3.3	7	4	6.7	4.1

Final solids contents are based on 1.5% solids in feed sludge, 10% oil solubles in sludge, and TC losses in the raffinate as shown. Lower final solids contents will increase the cost of the Carver Greenfield process, in most cases.

Reducing settling temperature from 175°F to 105°F will have the following overall and specific effects on heat and material balances:

- Eliminate the need for heating the feed to the 3rd concentration stage to 175°F (reduce both fuel and equipment costs).

- Reduce TC recycle loss (reduce cost).
- Eliminate the "superheat" in the concentrate feed to the evaporator (increase fuel costs).
- Increase weight of sludge solids to the incinerator, due to lower TC losses (reduce fuel costs).
- Reduce the solids content in the oil-sludge concentrate feed to the Carver-Greenfield evaporation step (effect on costs variable, depending upon specific plant size and feed concentration).

Combining all of the cost factors listed above, the net result is a considerable reduction in costs for the Esso process component, as shown in detail in Appendix D-7 and summarized in Table 42 below:

TABLE 42

PROJECTED COST SAVINGS FOR 105°F SETTLING: ESSO CONCENTRATION PROCESS COMPONENT

Sludge Load Tons/Day	Concentration Factor	Savings vs 175°F Settling - $/Ton	Revised Costs $/Ton
4.7	Max	8.4	39.7
	Min	9.5	44.4
14.2	Max	7.3	20.1
	Min	8.4	23.7
94.5	Max	6.4	11.1
	Min	7.5	14.3

Using the solids contents for the primary sludges as detailed previously, the solids contents for the combined primary + secondary sludges to Carver Greenfield process will range from 3.8-7.6%. The costs for these solids contents were obtained from the data in Figure 15 and are summarized below:

TABLE 43

COSTS FOR CARVER GREENFIELD PROCESS COMPONENT

PROJECTED DATA BASIS

Sludge Load Ton/Day	% Solids in Feed	Cost $/Ton
4.7	3.8	83
	7.6	74
14.2	3.8	41.2
	7.6	27.6
94.5	3.8	24.1
	7.6	11.5

The savings for the Carver-Greenfield process by operating under the projected data basis conditions can be estimated by comparing the operating costs for maximum and minimum feed concentrations for the two cost bases; these savings are tabulated below:

TABLE 44

COST SAVINGS FOR CARVER GREENFIELD PROCESS PROJECTED DATA BASIS

Plant Size Tons Sludge/Day	Savings in Costs - $/Ton for % Solids in Feed of Minimum (3.8%)	Savings in Costs - $/Ton for % Solids in Feed of Maximum (7.6%)
9.4	6	3
28.3	5	1
94.5	4	0.5
189	2	0.5

These values include the negative effect of operating at lower % solids contents compared to the original "present data" basis. The difference in solids contents at the maximum level is large (8.8 vs 7.6%) so the absolute magnitude of the net savings is relatively small.

Combining the projected costs for the two individual process components, costs for the projected data basis are shown in Figure 16 and tabulated below; thickening of primary sludge is included:

FIGURE 16

TREATMENT COSTS OF ESSO AND CARVER GREENFIELD PROCESS COMPONENTS-PROJECTED BASIS

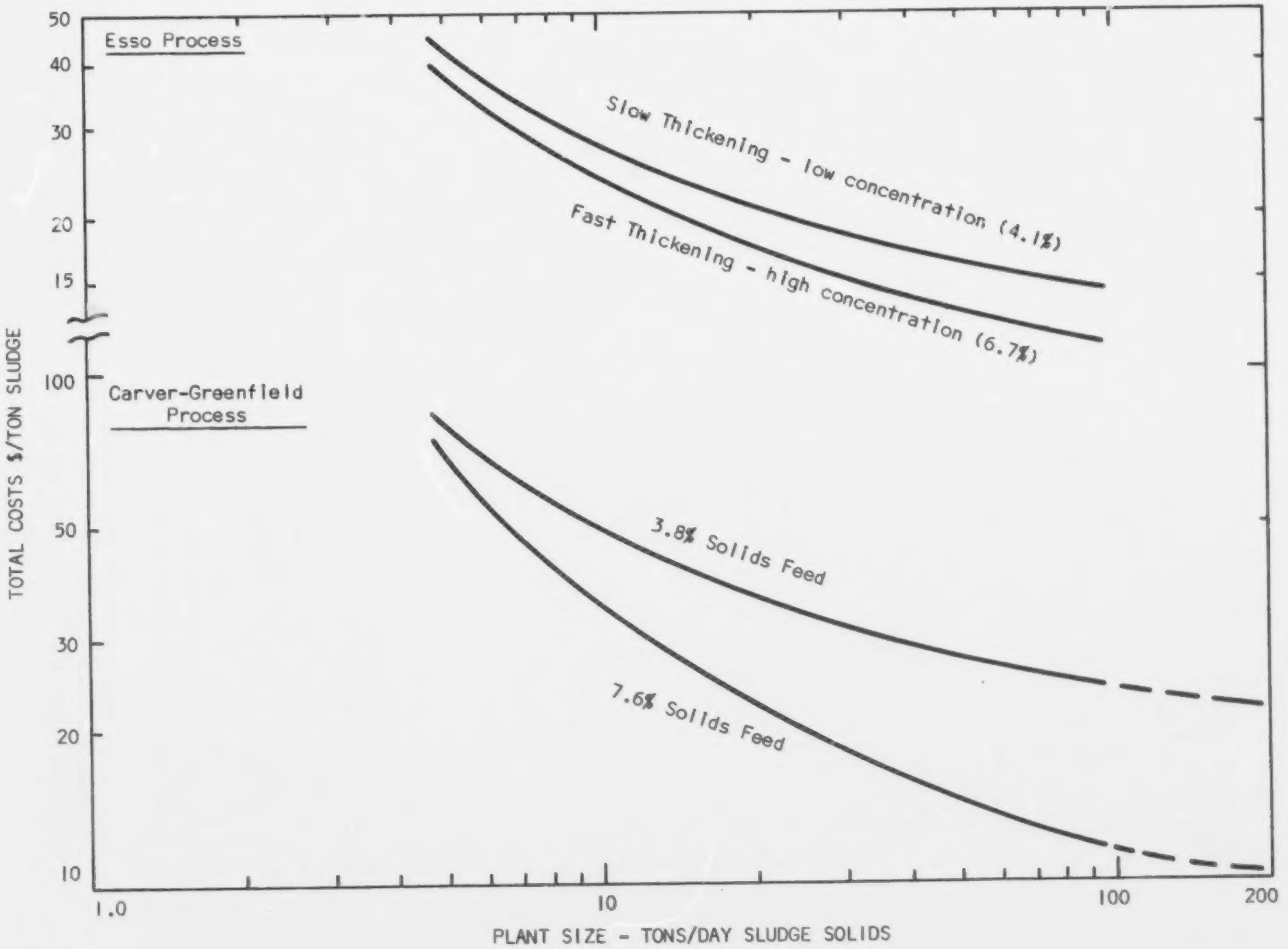

TABLE 45

TOTAL COSTS FOR 50/50 PRIMARY + SECONDARY SLUDGE PROJECTED DATA BASIS

Plant Size Sludge Loads Tons/Day	Treatment Cost Range, $/Ton		
	Esso	Carver Greenfield	Total Process
9.4	20-23	37-51	61-74
28.3	10-12	19-32	31-44
94.5	6.5-8.5	12-25	20-34
189	5.6-7	10.2-22	17-29

Comparing the costs for the original "present data" basis with the "projected data" basis, **average** savings of around $6/ton are expected for the combined Esso Carver Greenfield process:

TABLE 46

SAVINGS EXPECTED FOR LOW TEMPERATURE SETTLING, COMBINED ESSO-CARVER GREENFIELD PROCESS

Plant Size Sludge Load Tons/Day	Cost Savings $/Ton
9.4	1-6
28.3	3-8
94.5	3-8
189	4-10

Further cost reductions are also realistically possible, as discussed in subsequent sections.

7.5.2 50% Reduction in Oil-Sludge Settler Size Worth ∿$2/Ton

The oil sludge settlers for the 2nd and 3rd stages of the Esso concentration process were designed on the basis of the liquid depths measured in the pilot plant runs.

Reduction in the calculated size (surface area) of the oil-sludge settlers by 50% requires the use of a scale-up factor only 2x greater than the factor actually used. Scale-up from laboratory to pilot plant indicates that the factor actually used was very conservative, and that an increased factor would be a reasonable extrapolation. Cost savings calculated for a 50% reduction in 2nd and 3rd stage steelers only are detailed in Appendix D-10 and summarized below:

TABLE 47

POSSIBLE SAVINGS IN SETTLER SIZE

Secondary Sludge Load, Tons/Day	Cost Savings for 50% Area Reduction-$/Ton
4.7	2.3
14.2	1.9
94.5	1.6

This savings of ∿$2/ton can be added to the savings already projected for the 105°F settling, and increase the total cost reductions for all projected modification to $7-16/ton.

7.5.3 Alternative Process Modifications

Use of unthickened waste sludge, and adjusting sludge pH to 3.0 for the oil concentration step were two process alternatives that initially appeared attractive. These alternatives were considered in the cost analysis but both were found to be unacceptable, as detailed below.

7.5.3.1 Unthickened Sludge for Concentration Step

Use of waste sludge directly at 0.5% suspended solids content for the oil-concentration step would eliminate the considerable cost of the sludge prethickener. Other costs would be increased, however, due to the following factors:

- Added heating requirements to raise the sludge temperature to concentration temperature of 105°F.
- Added heating requirements to heat the oil-sludge mixture to 175°F for the final concentration stage.
- Increased settler size (surface area) due to the larger volumes of oil-sludge feed to each concentration stage.

The combined effect of these factors is a very large increase in costs compared to the use of a sludge thickener; specifics are summarized below:

- In winter, costs to preheat the sludge to the extraction temperature of 105°F were calculated at \$19.7/ton sludge, exclusive of the steam generation boiler; for prethickened sludge no added heat input is required.
- Increased fuel costs to heat the larger volume of oil sludge to 175°F from 105°F, for the 3rd stage of the concentration, is calculated at about \$2/ton.
- Total combined thickener + settler size (surface area) is greater for unthickened sludge than prethickened, even for the case of the slow settling waste sludge; use of prethickened sludge retains this advantage even if the projected 50% reduction in oil sludge settler requirements is assumed, as shown below:

Type Sludge to Esso Process	Total Thickener + Settler Area - ft^2	
	Present Design Basis	50% Reduction
Unthickened (0.5%)	2900	1450
Prethickened (1.5%)	1470	1135

7.5.3.2 Adjustment of Sludge pH to 3.0

Reducing the pH for extraction increased the concentration factor about 15-20%, and reduced TC losses in the raffinate by approximately the same amount. Depending upon the size of the plant and the solids content after concentration, savings fo up to ∿\$7/ton could be expected for secondary sludge.

On the debit side, however, this must include the cost of chemicals to acidify and to reneutralize, the cost of the chemical feeding and mixing equipment, and the cost of stainless steel for the 3rd stage settler and associated equipment. pH adjustment actually requires 3 separate steps: 1) oil-sludge adjustment from 5.5-6 to 3.0 before the final concentration step, 2) the adjustment of the aqueous raffinate back to a pH of ∿6, and 3) adjustment of the final concentration back to ∿6 prior to the Carver-Greenfield evaporation.

Sulfuric acid can be used for acidification; reneutralization requires the use of ammonium hydroxoxide (or ammonia) rather than lower cost caustic, in order to avoid a high sodium content in the feed to the incinerator. Chemicals + equipment for the pH adjustments are estimated at \$12-19/ton for the different size plants. These costs alone are considerably greater than the calculated savings for pH 3 operation, even without considering the large added cost for a stainless steel 3rd stage settler.

7.6 No Incentive for Sludge Prethickening to >1.5%

For a given sludge batch, increasing the solids content above 1.5% by further prethickening will not increase the final solids content attainable by the Esso concentration process; therefore, there will be not cost savings in the evaporation step.

There is no effect of higher solids on the heat balance, assuming use of the 105°F settling temperature. With 175°F settling, costs to heat the 3rd stage feed from 105°F-175°F will be somewhat lower due to the reduced volume of the oil-sludge mixture.

Pumping costs will be slightly lower with increased feed solids content, due to the reduced volumes of oil + sludge processes. TC losses will also be lower, since the raffinate volume will be reduced.

As noted above, the benefits of increased feed solids content are all minor, and will certainly be less than the increased costs for larger sludge thickeners.

7.7 Four Effect Evaporator System Reduces Fuel Costs

For the lower range of solids contents in the concentrates to the evaporation step, considerable fuel economy can be obtained using a 4 effect evaporator in place of a 3 effect system. As noted above, Carver-Greenfield normally restricts the use of 4 effect systems to $\leq$ 6% solids content and solids loads $\gtrsim$5 tons/day.

Using the Carver Greenfield design data for 4.5% and 6.0% solids (Appendix D-3), the potential savings (primarily in fuel), are about $5/ton as shown below for the projected data basis:

TABLE 48

INCENTIVES FOR 4 EFFECT EVAPORATION SYSTEM - PROJECTED DATA BASIS

Sludge Load Tons/Day	% Solids in Feed	of Effects	Total Investment $MM	Total Costs, $/Ton	Fuel Costs $/Ton
14.2	4.5	3	.835	38	4.45
		4	.822	33	(.16)*
	6.0	3		32	2.20
		4	.716	27	(3.04)*
94.5	4.5	3	2.8	20	4.45
		4	2.61	15	(.16)*
	6.0	3		15	2.20
		4	2.10	10	(3.04)*

*( ) Fuel equivalent of surplus heat generated.

The reduced fuel costs for solids contents of ∿4.5%, where the 4 effect system is essentially in thermal balance, represent real operating savings for all plants. Where surplus heat is generated, at solids contents >∿4.5%, some requirement for the steam for either heating or power must exist in order to credit the total reduced fuel cost as a "real" saving. The situation for each plant will have to be considered on an individual basis in order to establish the specific incentive. If a need for the surplus steam does not exist, a condenser system will have to be installed.

7.8 Thickener Costs May Be Greatly Reduced or Eliminated With Other Sludges

Based on a recent survey of plants by the EPA's Advanced Waste Treatment Research Lab (48), the typical waste activated sludge solids concentration was between 0.50 and 1.40% with a mean value of 1.0% solids.

The cost estimate for the Esso process component was based on a waste sludge concentration of 0.5% being thickened to 1.5%; therefore, for some sludges the thickener size requirements may be greatly reduced and in some cases even eliminated. The value of 1.5% solids after thickening was based primarily on heat balance considerations for the most severe conditions of mid winter operation.

At 1.5% solids and an oil/sludge ratio of 0.2 the oil-sludge concentrate can be heated to the required settling temperature in the barometric leg of the 3rd evaporation stage without additional heat input. Plants in warmer climates, where sludge temperatures as low as 40°F are not encountered, can operate at sludge concentrations below 1.5%; therefore thickener size requirements will be lower, even for 0.5% waste sludge feed, and particularly lower for waste sludges with solids contents above 0.5%.

The exact cost savings for reduced thickener size will depend upon the plant size, heat balalnce requirements, and sludge properties. Reductions in thickener area requirements of 25-100% for operation in warm climates and/or with waste sludges of high solids content appear likely for many plant locations. Cost savings of $0.3-8.6/ton sludge can be projected for these situations:

TABLE 49

POTENTIAL COST SAVINGS FOR REDUCED SLUDGE THICKENER AREA (1)

Plant Size Tons/Day Sludge	Thickening Rate	Cost Savings, $/Ton Sludge For Area Reduction of	
		25% (2)	100% (2)
4.72	Fast	1.1	4.5
	Slow	2.1	8.6
14.16	Fast	0.6	2.3
	Slow	1.0	4.1
94.5	Fast	0.3	1.1
	Slow	0.9	3.7

(1) Costs for secondary sludge only.
(2) Based on requirements for sludges tested.

7.9 Process Costs Expected to Be Lower for Other Sludges

As fully discussed in the section above, the total operating and investment costs are very sensitive to the process responses of the particular sludge batch processed; the specific responses of most concern are rate of thickening of waste sludge, and the final solids content achieved after the oil concentration step. Process economics were based on the range of responses for only three different sludge sources tested in one particular geographic area.

Most of the sludges tested can be characterized by slow thickening rates and low solids contents achieved by thickening; this is particularly the case for the Wards Island activated sludge. Other sludges in different parts of the country can be thickened to considerably higher solids contents than the samples used in this program (48). On average, the solids content after the oil concentration process was considerably higher for the sludges that thickened most rapidly (Bergen County) than for the slow thickening sludges, Wards Island. Sludges that thicken more rapidly and to higher solids content than Bergen County should therefore also produce higher solids concentrates from the Esso oil process.

The Wards Island sludges are recognized as being very difficult to dewater and so probably represent the high cost end of the spectrum for all plants. Costs for the Bergen County sludges are therefore probably more normal and representative of an "average" sludge, and should be used as the basis for comparative analysis with competetive processes.

7.10 Esso Carver Greenfield Process Costs Considerably Lower Than Current Commercial Processes

To better define the incentives for the new Esso Carver Greenfield process, a comparison was made with a process, proposed by the EPA for this purpose (48), which consists of the following sequence of steps:

- Primary sludge with 5% solids thickening to 8%.
- Waste activated sludge at 0.6% solids thickened to 4.5% by air flotation.
- Vacuum filtration of mixed thickened primary plus activated sludges, containing 5.7% solids, to a final solids content of 25%.
- Incineration of filter cake in a multiple hearth incinerator.

The operating cost breakdown for the process is tabulated in Table 50 below on the basis of the data provided by the EPA (48); the individual investment components, labor costs , maintanence, capital costs, etc. were derived from reference (37).

TABLE 50

COSTS OF CURRENT COMMERCIAL PROCESS

	Cost Effects of Plant Size (¢/1000 gal.)		
Process Component	1 MGD = .86/Day (1)	10 MGD = 8.6/Day	100 MGD 86/Day
Primary Sludge Thickening	1.466	0.218	0.076
Air Flotation, Activated Sludge(2)	2.276	1.005	0.753
Holding Tanks	1.005	0.263	0.097
Vacuum Filtration	8.389	5.147	3.698
Multiple Hearth Incineration	13.534	5.023	1.160
Totals, ¢/1000 gal	25.665	11.393	5.6807
$/Ton (3)	297.4	132.0	65.9

(1) 816.8 pounds of primary + 909.3 pounds of activated sludge/10^6 gallons.
(2) Includes chemicals cost of $1-2/ton.
(3) Mixed primary + activated sludge.

Comparing the costs for the representative present commercial process and for the Esso Carver Greenfield (ESSO-CG) process, the Esso-CG process has a considerable advantage even on the present data basis. The advantage is increased considerably on the projected basis.

TABLE 51

COST ADVANTAGE OF ESSO CARVER GREENFIELD PROCESS OVER CURRENT COMMERCIAL PROCESS

Plant Size Tons/Day Sludge		Cost, $/Ton Esso C-G	
	Commercial	Present Data	Projected
9.4	130	62-80	61-74
28.3	92	34-52	31-44
94.5	64	23-42	20-34
189	52	21-39	17-29

The advantage for the Esso C-G process becomes even more striking when the following factors not included in the above costs are considered:

- The upper limit for the cost range includes the effects of the poorest process response and use of unthickened primary sludge. The costs for thicker sludges will therefore range from the lower limit shown to about the midpoint of the total range.

- No credit was taken for anticipated cost reductions from reduced thickener and settler area requirements, which could total $1-5/ton of final mixed primary + activated sludge.

Considering all factors, the cost advantage of the Esso C-G process, for the above plant size range, is calculated to be ~$13-68/ton on a present data basis and ~$24-74/ton on a projected basis; the advantage for the Esso-Carver Greenfield process, on a percentage basis, actually increases with increasing plant size.

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10. APPENDICES

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TABLE A-1

A. Sample Processing (Sampling, Transport, Storage)

- The sludge was taken from the most appropriate sampling point at the sewage treatment plant with an open pail and poured into clean 2 and 5 gallon polyethylene wide mouth containers with bottom spigot valve.
- The filled sludge containers were packed in crushed ice for transport back to the GRL laboratory at Linden, New Jersey.
- Sludge was stored in a 40°F refirgerator after it was received in the laboratory, except for the short times required to remove material for analyses or experiments.
- Sludge samples for analyses as for testing were taken from the containers via the spigots after thorough mixing with a turbine agitator.
- To increase suspended solids content of the initial sludge sample as received, the contents of the container were allowed to settle to the desired degree and the supernate removed by careful siphoning.
- To determine suspended solids content duplicate 100cc samples were filtered thru Whatman #40 paper and dried to constant weight in accordance with the method given in "Standard Methods for the Examination of Water and Waste Water" (13th edition).

B. Equipment

- The oils used for the concentration step were heated to the desired temperature in 2 liter stainless steel pots, using resin kettle type heating mantles. Temperature control was provided by a West "Gardsman" controller, with the control thermocouple placed 1/4 of the distance up from the bottom of the pot in the outside surface (in contact with the heating mantle). The oil in the pot was agitated by means of a controlled nitrogen purge. Oil removal was made via a bottom drain line fitted with a needle valve.
- Sludge temperature was adjusted to the desired level by immersion in a standard constant temperature water bath.
- When using a propeller or turbine agitator, the mixing vessel was a 400cc plastic beater with two 1/4" steel baffles.
- Mixing by pump was carried out by premixing the sludge + oil in an agitated 1000cc resin kettle (with the bottom drain connected to the pump) for 10 seconds; the purpose of the premixing step was to insure a uniform oil/sludge ratio in the pump feed.

- for the laboratory program a single stage, open impeller Eastman Industries Co. pump was used; the pump was rated 6000 RPM, 1/20 HP.

- flow rate thru the pump was set at 10-15 seconds by use of needle valve on the 1/4" outlet line, set at 1/4" turn open for all runs.

- Settling of the oil-sludge mixture was carried out in either 250cc or 500cc straight sided dropping funnels with teflon plug stop codes; these settlers all had volume markings and were individually precaldrated.

- The settlers were held for the required time at the required temperature in constant temperature ovens of the recirculating forced air type.

C. Test Procedures

- Sludge and oil at the specified temperature were normally measured by volume in calibrated graduates, with the accuracy periodically checked by weight.

- The measured quantities of sludge and oil were added to the mixing vessel, then agitated for the required time using a stop watch.

- Immediately after mixing the mixture was charged to the settling vessel and placed into the constant temperature oven.

- The volumes of the separated oil-sludge and water raffinate phases were measured at the predetermined time intervals. The volume of any solids (sediment) which was not "captured" by the oil was also noted and the quantity of solids calculated from predetermined factors.

  - raffinate volumes were converted to density by applying the appropriate temperature **correction** in order to calculate concentration factor.

  - in many cases the raffinates were separated off periodically and weighed directly; this procedure was used for many of the runs involving TC determinations, and as a periodic check on the accuracy of the volumetric method.

TABLE A-2

ANALYSES OF ACTIVATED SLUDGES FROM SEWAGE PLANTS (1)

Test	L.F.-"B" 8/18/71 (3)	L.F.-"C" 9/7/71 (3)	L.F.-"D"	WI-"A" 9/20/71 (3)	WI-"B" 10/4/71 (3)
Suspended Solids, ppm	4600	4100	8200	6800	16,500
Volatile Solids as % of Suspended	64.1	64.6	68.8	70.1	69.6
Kjeldahl N, ppm	500	200		350	1,200
NH_4 N, ppm	90	28	37	9	120
NO_3N, ppm	<2	5	1	14	23
Total P_2O_5, ppm	11.7	22.7	24.1	11.5	5.5
Ortho P_2O_5, ppm	<1				
Acidity, ppm as $CaCO_3$	445	12		30	83
Alkalinity, ppm as $CaCO_3$	415	450		144	230
Total organic C, ppm(2)	29, 38, 37	47	51	36	
Total C, ppm(2)	45, 63, 56	99	61	9	
Respiration rate, ppm/O_2/min(4)			2.15	1.24	5.1
COD, ppm			9600	2700	7,500

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(1) Analyses by ERE analytical laboratory (AID).

(2) Run on Supernate; all other samples on total sludge.

(3) Date Sampled.

(4) LF = Bergen County, N.J.; WI = Wards Island, N.Y.

TABLE A-3

CHARACTERISTICS OF LITTLE FERRY, N.J. PLANT STREAMS

Plant Data(1)

- **Primary Treatment:**

Date		8/10	8/11	8/18	8/25
Raw Sewage	BOD_5	198			
	SS				
Effluent	BOD_5	138			
	SS		97	140	112
	COD			384	364
	Alkalinity		360		
	pH		6.6		

- **Secondary Treatment:**
  (Activated Sludge)

Date		8/10		8/11		8/18		8/25	
Mixed Liquor	Effluent								
	COD								
	SS				2080		1220		1340
	Alkalinity				340		1080		1380
Stabilizer tank	position	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Stabilizer tank	COD	105	96			5480	5040	3760	3720
	SS	5350	5070	5250	5360	5240	4490		
	pH	7.0	6.9	7.1	7.1				
	Alkalinity	700	660	500	420				
	NH_4N	19.7	13.5						

(1) Analyses as reported by plant laboratory; all values mg/liter = ppm.

TABLE A-4

WARDS ISLAND SEWAGE PLANT

AUG. 1971 PLANT ANALYSIS SUMMARY

• Raw Sewage	Max	Min	Average
- BOD Total Steam	159	50	97
- BOD Filtrate only	52	17	27
• Primary Effluent			
- BOD Total	97	28	64
- BOD Filtrate	40	16	29
• Return Sludge (Battery B, Step Aeration)(1)			
- Suspended Solids	6200	1700	4100
- % Volatiles in SS	80.2	72.0	75.2
• Final Effluent			
- Total Plant(2)			
• Suspended Solids	40	19	26
• BOD Total	28	10	18
• BOD Filtrate	14	5	9
- Battery B			
• Suspended Solids	16	7	11
• BOD Total	12	3	7
• BOD Filtrate	8	1	4

1) Source of sample for extraction tests.

2) Secondary treatment: ∼2/3 activated aeration, 1/3 step aeration on flow basis.

TABLE A-5

TRENTON, N. J. SEWAGE PLANT

JULY-AUG. 1971 PLANT ANALYSIS SUMMARY(1)

	July			August		
	Max	Min	Average	Max	Min	Average
Raw Sewage						
BOD	208	140	162	192	112	159
SS	524	186	350	602	272	382
Plant Effluent(2)						
BOD	55	38	48	52	23	39
SS	194	70	112	60	148	101

(1) All values ppm.

(2) From secondary treatment.

TABLE A-6

LABORATORY CONCENTRATION OF SECONDARY SLUDGES

(All runs with #4 Heating Oil, mixed by 1 pass thru pump)

Sludge Source	% Suspended Solids in Feed	Oil/Sludge Ratio	Settling Temperature°C	Settling Time-hrs	% Solids in Concentrate Uncorrected(1)
Wards Island "D" (3)	0.49	0.2	50	1	2.7
				5	3.7
				20	4.5
		0.2	80	1	3.3
				5	7.1
				20	7.8
	1.7	0.2	50	1	2.3
				5	3.8
				20	6.1
		0.2	80	1	3.1
				5	4.6
				20	6.7
Bergen County (L.F.)-"I"(4)	0.52	0.2	40	2	4.9
				4	5.7
				18	7.7
		0.2	60	2	4.2
				4	5.2
				18	6.9
		0.2	80	2	6.1
				4	9.5
				18	9.5
	2.1	0.2	40	2	2.9
				4	3.6
				18	5.4
			60	2	3.4
				4	4.2
				18	7.5
			80	2	4.4
				4	5.5
				18	7.1

(Continued)

Sludge Source	% Suspended Solids in Feed	Oil/Sludge Ratio	Settling Temperature	Settling Time-hrs	% Solids in Concentrate Uncorrected(1)
Trenton "B" ( )	1.24	0.1	80	1	4.7
				2.5	6.1
				17	6.1
		0.2	60	1	3.0
				2.5	4.6
				17	5.1
		0.2	80	1	4.0
				2.5	6.8
				17	6.8
		0.4	80	1	3.8
				2.5	5.1
				1.7	7.7

(1) Calculated from feed solids x concentration factor; value reported previously.
(2) Activated.
(3) Trickle Filter.

TABLE A-7

LABORATORY CONCENTRATION OF MIXED PRIMARY & SECONDARY SLUDGES

Sludge Source	% Suspended Solids in Feed	Oil/Sludge Ratio	Settling Temperature	% Solids in Concentrate Time-hrs	Uncorrected (1)
Wards Island "D"	2.1	.2	50	1	2.9
				5	4.2
				20	5.5
		.2	80	1	3.6
				5	7.2
				20	11
		.4	80	1	3.7
				5	8.4
				20	13
Bergen County (L.F.)-"I"	1.7	.25	70°	1	3.7
				7	5.2
				20	5.2
	3.5	.25	50	1	3.7
				7	5.9
				20	7.1
		.25	80	1	4.0
				7	6.6
				20	7.1
		.4	50	1	3.5
				7	4.7
				20	6.6
		.4	80	1	3.9
				7	6.0
				20	7.1

(1) Calculated from feed solids x concentration factor; value reported previously.

TABLE A-8

EFFECT OF SETTLING TEMPERATURE ON CONCENTRATION FACTOR

Sludge Batch	% Solids in Feed	Oil	Agitation Type	Concentration Factor													
				25°C		40°C		45°C		50°C		60°C		80°C		95°C	
				1 hr	17/21	1 hr	17/21	1 hr	17/21	1 hr	17/21	1 hr	17/21	1 hr	17/21	1 hr	17/21
Secondary Sludges																	
Bergen County A	0.86	#4 HO	hand	2.0	3.1												
		#1 Varsol	hand	2.0	2.8							2.7	5.0				
	2.3	#4 HO	Waring Blender	1.4	2.1							2.3	5.0				
		#1 Varsol	Waring Blender	1.6	2.8							1.6	3.0				
												2.0	3.3				
Bergen County B	0.89	#2 HO	Waring Blender	1.6	2.5			1.6	3.8			2.2	4.5				
			propeller	1.8	2.6			1.8	3.3			2.0	3.3				
	0.89	LOPS	Waring Blender	1.7	2.5			1.8	4.2			2.2	5.0				
			propeller	1.9	2.7			2.0	4.1			2.2	4.5				
	1.7	#2 HO	Waring Blender	1.6	2.3			1.7	3.3			1.9	3.3				
		LOPS	propeller	1.6	2.3			1.8	3.3			1.8	3.3				
Bergen County E	1.0	#4 HO	pump											3.1	6.8	3.2	1.0
	1.8	#4 HO	pump											2.0	3.5	2.2	3.6
Bergen County I	0.52	#4 HO	pump			8.1	13.3					9.5	14.9	11.8	18.3		
	2.1		pump			1.4	2.6					1.6	3.4	2.1	3.7		
Bergen County J	0.7	#4 HO	pump			4.3	8.5					4.0	9.8	5.8	16.0		
	3.0	#4 HO	pump			1.2	2.2					1.2	2.4	1.4	3.0		
Bergen County K	0.5	#4 HO	pump			6.4	12.9					11.8	18.4				
	2.2	#4 HO	pump			2.0	4.0					3.0	4.4				
Wards Island D	0.59	#4 HO	pump							5.5	9.2			6.8	15.9		
	1.75	#4 HO	pump							1.3	3.5			1.8	3.8		
Trenton B	1.2	#4 HO	pump									2.9	4.1	3.2	5.5		
Primary + Activated Sludges																	
Bergen County I	3.5	#4 HO	pump							1.1	2.1			1.2	2.1		
Wards Island D	2.1	#4 HO	pump	1.4	2.7									1.9	5.2		

TABLE A-9

TURBINE AGITATOR NOT EFFECTIVE FOR CONSISTENTLY HIGH SOLIDS CAPTURE

Activated Sludge Batch	% Suspended Solids	Agitator Type	Mixing Time Seconds	Oil	Oil/Sludge	Solids Capture - %
Bergen County "C"	0.40	Turbine - 350 RPM	30	#4 HO	0.3	< 75
↓	0.74	↓		#4 HO	0.1	< 75
	0.74			#4 HO	0.3	93
	0.74			LOPS	↓	< 75
	1.2			#4 HO		95
	1.2			#4 Varsol		92
	2.1			#4 HO		98
	2.1			#4 Varsol		95
Bergen County "D"	.82	Turbine - 350 RPM	20	#4 HO	0.2	< 75
	.82	Waring Blender	5	#4 HO	0.2	~ 99
	.82	Turbine - 500 RPM	20	#4 HO	0.2	90
	2.3	Turbine - 500 RPM	20	#4 HO	0.2	97
Wards Island "B"	0.55	Turbine - 350 RPM	20	#4 HO	0.2	< 50
	0.55	Waring Blender	2	#4 HO	0.2	99
	1.65	Turbine - 350 RPM	20	#4 HO	0.2	91
	1.65	Turbine - 350 RPM	60	#4 HO	0.2	94
	1.65	Waring Blender	2	#4 HO	0.2	99

TABLE A-10

CENTRIFUGAL PUMP EFFECTIVE MIXER; BERGEN COUNTY SLUDGES - BATCH "E"

Age, Days (4)	Type Sludge	% Susp. Solids	Oil	Oil/Sludge Ratio	Agitation	Settling Temp. °C	Concentration Factors after Settling 0.5 hr	1 hr.	2/2.5	3/3.5	4/4.5	5/6	18/20	64/68	Approx. Solids Capture, %
0	activated	1.0	#4 HO	0.2	20 sec./500 RPM turbine	80			6.6						75
	"	"	"	"	" " (1)	"			4.5	6.7		6.7	(8.5)	10	90
	"	1.8	"	"	" "	"		(1.6)	1.9	2.0	2.2		(3.2)	3.5	98
	"	1.0	"	"	2 passes through pump (2)	"		3.2	3.5	4.0			(5.4)	6.3	99
	"	"	"	"	40 sec./680 RPM Eppenbach	"		3.5	4.6	5.0			(6.4)	7.1	99
	mixed primary + activated	1.9	"	"	20 sec./570 RPM turbine	"		2.1		2.7		3.2	(3.8)	4.5	95
		"	"	0.4	" "	"									70
		"	#1 Varsol	0.2	" "	"									60
	"	"	#4 HO	"	5 sec. W. Blender	"		2.9	3.2	3.5	3.7		4.4	5.9	99
4	activated	1.0	"	"	1 pass through pump (2)	95	2.9	3.5	4.1	5.6	6.1	6.2	10		"
	"	"	"	"	" " (1)	"	3.0	3.5	4.6	5.2	5.6	5.9	9.0		"
	"	1.8	"	0.1	" "	80	1.8	2.2	3.2	3.7	4.2		5.2		"
				0.2	" "	"	2.0	2.8	3.2	3.6	3.9		4.8		"
				0.4	" "	"	1.9	2.8	3.1	3.2			4.5		"
				0.2	" "	95	2.3	2.8		3.5	4.1		5.0		"
	mixed primary + activated	1.9	"	0.1	" "	80	1.8	2.4	2.8	3.7			5.0		"(3)
				0.2	" "	"	2.4	2.8	3.2	3.7			4.5		"
				0.4	" "	"	1.7	2.2	2.4	3.1			3.8		
5	activated	1.0	"	0.2	" "	"		3.3	4.1	4.3	4.8		8.0		"
	"	"	"		2 passes "	"		3.3	3.5	4.2	4.3		5.6		"
	"	"	#1 Varsol	"	5 sec./680 RPM Eppenbach	"		(3.3)	3.6				6.1		"
	"	1.8	#4 HO	"	1 pass through pump (2)	"		1.5	1.8	2.0			2.8		"
	"	"	"	"	" "	95		1.7	2.0	2.1	2.4		3.0		"
	"	"	"	"	" "	80		1.3	1.7	1.8			2.2		"
	"	"	LOPS	"	" "	95		1.4	1.7	1.9			2.6		"
	"	"	#4 HO	"	2 passes	95		1.9	2.3				2.5		"
	"	"	"	"	5 sec./680 RPM Eppenbach	80		1.2	1.5				2.5		"

---

(1) For this run sludge put through centrifugal pump twice before extraction.

(2) Sludge + oil premixed with turbine before pump mixing.

(3) Solids not as firmly held in oil layer as for higher oil/sludge ratios.

(4) Days stored at 40°F after sampling from plant.

TABLE A-11

OIL CONCENTRATION PROCESS WORKS FOR DIFFERENT TYPE SLUDGES: BERGEN COUNTY BATCH "D"

Sludge Age, Days(1)	Type	Feed % Susp. Solids	Oil	Oil/Sludge Ratio	Agitation	Batch Size, cc	Settling Temp, °C	Concentration Factor after Settling (1) 1 hr.	2 hrs.	3 hrs.	4 hrs.	19/22 hrs.	72 hrs.	Approx. Solids Capture, %
0	activated	.82	#4 HO	.2	5 sec. W. Blender	200	80	6.7	7.4		10.0	10.0	10.0	99
	"	"	#1 Varsol	"	"	"	25	2.3	3.6		4.5	5.7	5.7	
	"	2.3	#4 HO	"	"	"	25	2.9	3.3		3.6	4.5	4.5	
	"	"	#1 Varsol	"	"	"	80	2.6	3.1		3.1	4.4	5.0	
	"	.82	#1 Varsol	"	20 sec./500 RPM turbine	"	"	4.0	4.0			5.0		
	"	2.3	#4 HO	"	"	"	"	2.6	2.6			3.3		
4	digested	2.8	#4 HO	.1	20 sec./500RPM turbine	"	"	1.2	1.8	2.1		2.1		65
	"	"	"	"	2 sec. W. Blender	"	"	1.6	1.8	2.0		2.2		
	"	"	"	.3	5 sec. W. Blender	"	"	1.3	1.5	1.6		1.8		
	"	"	#1 Varsol	.3	20 sec./500 RPM turbine	"	"	1.3	1.6	1.7		2.5		92
	"	"	"	.3	2 sec. W. Blender	"	"	1.3	1.6	1.7		2.4		94
	"	"	"	.1	5 sec. W. Blender	"	"	1.6	1.8	1.9		2.9		
5	activated	2.3	#4 HO	.2	10 sec./500 RPM turbine	"	"	1.3	1.6	1.9		2.9		98
	"	"	"	"	20 sec./500 RPM turbine	"	"	1.3	1.6	1.9		2.7		
	"	"	"	"	50 sec./500 RPM "	"	"	1.2	1.6	1.8		2.9		
	primary + activated	2.7	#4 HO	.1	5 sec. W. Blender	"	"	2.5	3.1		3.3	4.4		99
		"	"	.2	"	"	"	2.7	3.3		3.6	5.0		
		"	#1 Varsol	.2	"	"	"	2.7	3.3		3.6	4.8		
		4.5	#4 HO	.2	"	"	"	1.4	1.7		1.9	2.5		
		"	"	.2	20 sec./500 RPM turbine	"	"	1.2	1.3		1.7	2.2		
		"	"	.4	"	"	"	1.3	1.6		1.8	2.4		
6	activated	2.3	#4 HO	.2	20 sec./500 RPM, 1 turbine	500	"	1.5	1.8	2.0				
	"	2.3	"	.2	" 2 turbines	750	"	1.4	1.7	1.9		3.3		
	primary + activated	4.5	"	.2	" 1 turbine	450	"	1.2	1.3	1.4		2.0		
7	activated	.82	#4 HO	.2	20 sec./500 RPM turbine	200	25	2.8		3.3	3.3	(3.5)	3.6	90
	"	2.3	"	.2	5 sec. W. Blender	"	80	2.5	2.9	3.2	3.4	(4.4)	4.6	99
	"	"	"	.2	20 sec./500 RPM turbine	"	"	2.0		2.3	2.6	(3.2)	3.6	97

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(1) Data not corrected for TG losses in raffinate and oil content of feed sludges; the estimated correction factors for these items are 0.85 + 0.90, respectively.

TABLE A-12

EFFECT OF SOLIDS CONTENT OF FEED ON FINAL SOLIDS CONCENTRATION 80°C SETTLING

Sludge Batch(1)	Sludge Type	% Suspended Solids in Feed	Type Mixing	Concentration Factor(2) 1 Hr	Concentration Factor(2) 2 Hr	Concentration Factor(2) 20 Hr	% Solids in Concentrate 1 Hr	% Solids in Concentrate 2 Hr	% Solids in Concentrate 20 Hr
LF-D	activated	0.82	Waring Blender	6.7		10	5.5		8.2
		2.3		2.6		4.5	5.7		8.6
LF-E	activated	1.0	pump	3.3		8.0	3.3		8.0
		1.8		2.8		4.8	5.0		8.6
LF-E	primary +	1.9	Waring Blender	2.9		4.4	5.5		8.3
LF-D	activated	2.7		2.5		4.5	6.8		10.8
		4.5		1.2		2.1	5.4		9.5
LF-K	activated	0.52	pump		11.8	18.4		5.9	9.2
		0.96			6.4	10.4		6.3	10.3
		2.1			3.0	4.4		6.6	9.7
LF-J	activated	0.70	pump	6.0		15.0	4.2		10.5
		0.75		6.7		12.8	5.0		9.6
		1.0		5.4		10.5	5.4		10.0
		1.5		3.5		6.9	5.3		10.3
		3.0		1.45		3.4	4.4		10.1
		3.0		1.40		3.6	4.2		10.7
WI-B	activated	0.55	Waring Blender	4.5		8.2	2.5		4.6
		1.6		1.9		3.5	3.0		5.2
		2.3		1.3		2.0	3.0		4.6
WI-D	activated	0.59	pump	6.8		15.9	3.3		7.8
		1.75		1.8		3.9	3.1		6.9

(1) LF = Bergen County, WI = Wards Island

(2) All runs wth oil/sludge ratio = 0.2, 80°C settling temperature, #4 Heating Oil.

TABLE A-13

EFFECT OF INITIAL FEED SOLIDS CONTENT AND OIL/SLUDGE RATIO ON FINAL OIL/SLUDGE RATIO

% Suspended Solids in Feed Sludge		Oil/Sludge Ratio for Concentration		Oil/Solids Weight Ratio (#/#)	
Total	Basis (1)	Vol.	Wt. (2)	Uncorrected	Corrected (3)
0.50	0.45	0.05	.04	8.9	10.5
		0.1	.08	17.8	21.0
		0.2	.16	35.6	42.0
		0.3	.24	71.2	65.0
1.0	0.90	0.1	.08	8.9	10.5
		0.2	.16	17.8	21.0
		0.3	.24	26.7	32.5
1.5	1.35	0.1	.08	6.0	7.0
		0.2	.16	11.9	14.0
		0.3	.24	17.9	21.0
2.0	1.80	0.1	.08	4.5	5.2
		0.2	.16	8.9	10.5
		0.3	.24	13.4	16.2

(1) Assume 10% oil solubles in sludge.
(2) Assume oil specific gravity of 0.8.
(3) For 15% TC loss during concentration step.

TABLE A-14

EFFECT OF OIL SLUDGE RATIO ON CONCENTRATION FACTOR

Sludge Batch	Sludge Type	% Solids in Feed	Oil Used	Settling Temperature	Type Mixing	Oil/Sludge Ratio	Concentration Factor	
							1 Hr	20 Hr
WI-B	activated	0.55	#1 Varsol	80	Waring Blender	.1	4.5	8.1
						.2	2.9	6.7
						.4	2.8	6.0
			#4 H.O.			.1	6.4	11.2
						.2	4.5	8.1
						.4	4.2	8.1
		1.65	#4 H.O.			.1	2.0	2.9
						.2	2.1	3.3
						.4	1.8	3.3
		1.65	#1 Varsol			.1	1.7	4.2
						.4	1.6	3.7
		2.35	#4 H.O.			.1	1.3	1.7
						.2	1.3	2.0
						.4	1.0	1.7
		2.35	#1 Varsol			.1	1.2	2.3
						.2	1.1	2.0
LF-D	primary + activated	2.7	#4 H.O.	80		.1	2.4	4.0
						.2	2.6	4.5
LF-E	activated	1.9	#4 H.O.	80	pump	.1	2.4	5.1(1)
						.2	2.8	4.5
						.4	2.2	3.8
	activated	1.9	#4 H.O.			.1	2.3	5.2(1)
						.2	2.9	4.8
						.4	2.8	4.5

LF-J	activated	3.0	#4 H.O.	80	pump	.2	1.4	3.4
						.33	1.5	3.1
Trenton B	trickle filter	1.24	#4 H.O.	80	pump	.1	3.5	4.9(1)
						.2	3.2	5.5
						.3	3.1	6.2

Average Values (2)

Oil/Sludge Ratio	Concentration Factor	
	1 HR	20 HR
.1	2.8	5.0
.2	2.5	4.6
.3	2.2	4.4

(1) Oil + sludge-water interface less stable than at higher o/s ratios.

(2) Based on direct comparisons only.

TABLE A-15

PROPERTIES OF POTENTIAL OILS FOR SLUDGE DEWATERING

Oil	Specific Gravity 60°F	Composition, Vol. %					Distillation, °F			Flash Point, °F	Viscosity	Approximate Cost, $/gal
		Paraffins	Napthenes	Aromatics		Olefins	IBP	50%	Dry			
				Total	C_8t							
#2 Heating Oil	.87	38	30(1)	32			335	499	648	158	2.3cs at 100°F	.115
#4 Heating Oil	.884						342	577	860	200	50SSU at 100°F	.110
Varsol #1	.789	46.1	39.8	14.0	13.0	0.1	319	342	380	105	.92cp at 25°C	.19
Varsol #4	792	54.5	31.5	13.8	13.8	0.2	363	373	402	140	1.15cp at 250°C	.20
LOPS(2)	.796	54.3	43.3	2.4			383	426	474	152	1.2cs at 100°F	.20
ISOPAR M	.782	79.9	19.7	0.3	0.3	0.1	405	434	484	172	2.43cp at 25°C	.31
CORAY 37	.901	69.5		30.5						310	80SSU at 100°F	.19

(1) Napthenes + Olefins.
(2) Low Odor Paraffin Solvent (Product of Enjay Chemical Co.)

TABLE A-16

COMPARISON OF OILS FOR CONCENTRATION PROCESS

Sludge Batch	% Solids in Feed	Oil/Sludge Ratio	Settling Temperature	Type Mixing	Oil Used	Concentration Factor 1 Hr	Concentration Factor 20 Hr
W.I-A	1.5	0.4	80	Turbine	#4 H.O.	1.6	2.3
					LOPS	1.5	2.1
W.I-B	0.55	0.2	80	Waring Blender	#4 H.O.	4.5	8.1
					#1 Varsol	2.9	6.7
	1.65	0.2	"	"	#4 H.O.	2.1	3.3
					#1 Varsol	1.7	4.0
	1.65	0.6	"	"	#4 H.O.	1.4	3.0
					#1 Varsol	1.4	3.0
					#2 H.O.	1.3	2.7
	2.35	0.2	"	"	#4 H.O.	1.3	2.0
					#1 Varsol	1.1	2.0
LF-D	0.82	0.2	"	"	#4 H.O.	5.8	8.2
					Coray 37	3.8	8.2
	2.3	0.2			#4 H.O.	2.5	4.2
					Coray 37	1.5	2.7
					#1 Varsol	2.6	4.6
LF-E	1.88	0.2	"	"	#4 H.O.	1.5	2.8
					#1 Varsol	1.3	2.3
					LOPS	1.4	2.6
LF-A	2.3	0.2	60	"	#4 H.O.	1.7	2.9
					#1 Varsol	2.0	3.3

LF-B	0.9	0.2	60	Propeller	#2 H.O.	2.2	4.2
					LOPS	2.3	3.8
	1.7	0.2	"	"	#4 H.O.	1.5	4.0
					#2 H.O.	1.2	3.5
					LOPS	1.9	4.2

Averages:

Oil	Number of Tests in Direct Comparison	Concentration Factor 1 Hr	Concentration Factor 20 Hrs
#4 H.O.	7	2.1	3.8
#1 Varsol		1.9	3.7
#4 H.O.	3	1.5	3.1
LOPS		1.6	3.0
#2 H.O.	2	1.7	3.8
LOPS		2.1	4.0
#4 H.O.	2	1.5	3.5
#2 H.O.		1.3	3.1

TABLE A-17

EFFECT OF SURFACTANTS - WARDS ISLAND BATCH A

Oil: #4 Heating Oil. Oil/Sludge Ratio = :2 Suspended Solids in Feed: 1.5%

Settling Temperature: 25°C for 1 - 5 hrs, Then 80°C for 5 - 22 hrs.

Mixing: 60 seconds, 350 RPM Turbine

	Concentration Factors(1)		
Surfactant(2)	1 Hr.	5 Hrs.	22 Hrs.
None	1.2 - 10	1.5	3.1
1% Triton X-15	1.2 - Trace	1.5	3.3
5% Triton X-15	1.3 - 1	1.5	3.3
1% Tallene	1.3 - 3	1.6	3.1
1% Armeen T	1.3 - Tr.	1.5	3.4
1% Span 85	1.2 - Tr.	1.4	3.1
2% Armoflo 49(3)	1.3 - Tr.	1.6	3.1
1% Atmos 300	1.2 - Tr.	1.5	3.1
1% Paranox 24	1.1 - Tr.	1.5	3.3
1% Oleic Acid	1.2 - 5	1.6	3.1

(1) First number = concentration factor, 2nd value = cc sediment; 10 cc ≅ 5% of feed solids.

(2) Surfactant dosage based on oil; 1% dosage in oil ≅ 0.1# Surfactant/1.0# Sludge Solids.

(3) 1% active amine.

TABLE A-18

EFFECT OF SLUDGE pH ON CONCENTRATION (pH 4)

Sludge(1)	% Susp Solids	Oil	Concentration Factor(3)			
			pH 6.5 - 7(4)		pH 4.0	
			1 hr.	16/20 hrs.	1 hr.	16/20 hrs.
L.F.-B	1.7	#4 HO	1.6	3.6	2.2	3.8
		#1 Varsol	1.5	2.9	1.8	2.9
L.F.-C	2.1	#4 HO	1.9(8)	6.0	3.0(10)	6.0
		#4 HO	2.0		2.3	
		LOPS + Varsol	2.1		2.5	
W.I.-A	1.5	#4 HO	1.6	2.3	1.9	3.7
		LOPS	1.5	2.1	1.9	3.3
W.I.-A	1.5	#4 HO	2.6(17)		2.6(3)	
		#4 Varsol	3.0(37)		4.0(22)	
W.I.-A	1.5	#4 HO	2.3(30)		2.3(10)	
		#4 Varsol	2.3(5)		2.3(10)	
W.I.-A	1.5	#4HO	1.9(10)	3.7	1.6(5)	3.3
		Average	2.0(18)	3.6	2.4(10)	3.9

---

(1) L.F. = Bergen County, W.I. = Wards Island.

(2) Averaged results for 1/2 factorial.

(3) Number in parenthesis cc sediment for 150 cc feed; 10 cc sediment ≃ 7% of initial feed solids; where no sediment shown, value < 1 cc.

(4) pH of sludge before acidification.

TABLE A-19

EFFECT OF INITIAL SLUDGE pH ON CONCENTRATION (pH 3)

#4 Heating Oil, pump mixing, 0.2 Oil/Sludge ratio

Sludge Batch	% Suspended Solids in Feed	Settling Temp. °C	Settling Time-hrs	% Solids in Concentrate pH 6.7	% Solids in Concentrate pH 3.0	Solids in Concentrate Ratio pH 3.0/6.7
Wards Island - D	0.49	80	1	3.3	5.6	1.7
			5	7.2	9.6	1.33
			21	7.8	9.6	1.23
	0.49(1)	80	1	3.6	4.4	1.23
			5	7.8	8.2	1.05
			21	8	8.2	1.0
	1.7	80	1	3.1	4.0	1.29
			5	4.9	7.4	1.50
			21	6.7	7.7	1.15
Bergen County - K	2.2	40	1	3.9	4.1	1.06
		40	5	6.4	7.3	1.14
		80(2)	21	9.9	12.2	1.23

Improvement with pH 3 sludge vs pH 6.7

Ave. pH 3/pH 6.7:		
	1 Hr.	1.33
	5 Hr.	1.25
	21 Hr.	1.15

(1) Repeat run next day.
(2) Settling temperature raised to 80°C after 5 hours.

TABLE A-20

TC LOSSES IN RAFFINATE

- W.I. - "D" Activated Sludge(3)

% Susp. Solids	Special Conditions	Settling Temp. °C	Settling Time Hrs.	Total(2) Carbon PPM	Feed C in Raffinate-% (4)
0.49		50	2	78	1.3
			6	142	5.1
			20	240	9.0
		80	2	174	6.5
			6	340	15.0
			20	570	20.5
	pH 3.0	80	2	306	14.2
			6	325	17.5
			20	310	15.7
	H_2O removal after 2, 6 hrs	80	2	226	9.7
			6	1140	12.1
			20	2310	15.5
1.75		50	2	126	0.4
			6	370	3.0
			20	1230	11.8
		80	2	813	7.0
			5	1035	9.9
			19	1200	13.4
	pH 3.0 H_2O removal at 4 hrs.	80	2	545	5.3
			20	2390	10.1
	H_2O removal after 5 hrs	50	2	285	1.4
			5	650	5.1
			19	2600	12.4

(1) In sludge feed.

(2) In raffinate.

(3) All runs with pump mixing, #4 H.O., oil/sludge ratio = 0.2

(4) Based on total carbon in sludge solids.

TABLE A-20 (Continued)

TC LOSSES IN RAFFINATE

- LF-I Activated Sludge(3)

% Susp. Solids	Settling Temp. °C	Settling Time Hrs.	PPM Total Carbon	Feed C in Raffinate - %
0.52	40	2	340	10.1
		18	390	13.1
	60	2	570	12.0
		18	630	26.2
	80	2	440	16.2
		18	578	24.1
2.1	40	2	550	1.8
		18	1500	8.5
	60	2	1250	6.1
		18	1265	17.8
	80	2	1470	10.2
		18	2022	19.2

TABLE A-20 (Continued)

- LF-K Activated Sludge(3)

% Susp. Solids	Special Conditions	Settling Temp. °C	Settling Time Hrs.	Feed C in Raffinate - %
2.2	staged temperature settling	40	1.7	4.8
		40	5.8	6.4
		80	7.8	8.3
		80	21	11.5
	pH adjusted to 3.0 staged temperature settling	40	2	4.8
		40	5.5	5.7
		80	7.5	7.0
		80	20.5	9.7
		40	1.8	4.9
			5.8	6.3
			21	8.0
		80	1.8	8.1
			5.8	11.5
			21	13.9

TABLE A-21

TC LOSSES IN RAFFINATE

- LF-G Activated Sludge (No Oil Controls)

% Solids in Feed	Oil	Mixing	Temp. °C	Time Hrs.	PPM Total Carbon	% Feed C in Raffinate	% Feed C Solubilized
0.55	none	Waring Blender	80	1	410	17.4	22
				2.8	730	34.7	40
				21	950	48.7	52
0.36	none	pump	25	1	52	3.3	4.2
				3	135	10.0	11.4
				21	220	16.6	18.0
1.82	none	no agitation	25	1	255		4.3
				3	285	(1)	4.7
				21	535		8.0
1.82	none	pump	25	1	310		5.2
				3	400	(1)	6.7
				21	910		15.2

(1) Solids did not separate to give raffinate phase.

Wards Island B - Activated Sludge

% Susp. Solids	Oil	Mixing	Settling		PPM Total Carbon	% Feed C in Raffinate
			Temp. °C	Time Hrs.		
0.55	#4 H.O.	Waring Blender	80	20	410	18.5
	#1 Varsol	Waring Blender	80	20	480	18.5
1.65	#4 H.O.	Waring Blender	80	7	730	6
				20	1090	11
	#1 Varsol			7	640	5
				20	1430	12
2.35	#4 H.O.		80	1	920	1.9
	#1 Varsol			1	1010	2.1
	#1 Varsol			20	2180	15.5

Wards Island A - Activated Sludge

1.5	#4 H.O.	turbine	80	3	480	5.8
	#2 H.O.	turbine	80	3	460	5.6
	#1 Varsol	turbine	80	3	260	3.1
	#4 Varsol	turbine	80	3	260	3.1
	#4 H.O.	turbine	80	20	710	9.5
	LOPS	turbine	80	20	1180	12.9

TABLE A-21 (Continued)

- LF-D Activated Sludge

% Susp. Solids	Oil	Mixing	Settling Temp. °C	Settling Time Hrs.	PPM Total Carbon	Feed C in Raffinate %	Feed Solubilized %
0.80	none	Waring Blender	80	1	260	6.9	8.6
				22	540	16.1	18.4
2.3	none	turbine	80	1	815	4.6	10.9
				22	1480	13.5	19.7
none(1)	#4HO	turbine	80	1	85	--	
				22	43	--	
none(1)	#1 Varsol	Waring Blender	80	1	11	--	
				22	16	--	
.80	#1 Varsol	Waring Blender	25	1	95	2.5	3.4
				22	180	4.3	6.1
.80	#4HO	Waring Blender	80	1	225	6.4	7.6
				22	305	9.1	11.1
2.3	#4HO	Waring Blender	25	1	190	1.7	3.5
				22	415	4.3	5.6
2.3	#1 Varsol	turbine	25	1	53	0.2	0.3
				20	220	0.9	2.9
2.3	#4HO	turbine	80	1	350	3.9	4.6
				20	1020	10.8	13.9
2.3	#1 Varsol	Waring Blender	80	1	325	2.7	4.3
				22	860	8.8	10.4

(1) Supernate with solids removed by filtration.

TABLE A-22

ANALYSIS OF CORAY 37 RECYCLE OIL FROM HERSHEY, PA.

ESSO RESEARCH AND ENGINEERING COMPANY
ANALYTICAL AND INFORMATION DIVISION
P. O. BOX 121, LINDEN, N. J. 07036

J. W. HARRISON
DIRECTOR

November 17, 1971

Dr. T. M. Rosenblatt
Government Research Laboratory
Building #1
Esso Research Center

Dear Ted:

Attached is a brief interpretation of the IR spectra of the Coray oil used in your extraction studies. The bulk of the material in the oil is a soap, probably calcium stearate. The spectra will be kept on file for future use.

If I can be of further help, please call.

Very truly yours,

J. J. ELLIOTT

JJE/bam

Attachment

cc: Messrs. R. E. Barnum
R. A. Brown

TABLE A-22 (Continued)

ATTACHMENT

- IR of used oil. Coray used in the reference beam.
    - Organic acids (1715 cm^{-1})
    - Organic esters (1748 cm^{-1})
    - Soap (intense peak at c. 1570 cm^{-1})
    - Strong broad band at 1100 cm^{-1} could be a C-O-C bond. Not hydroxyl, for 3300 cm^{-1} region only has a relatively weak peak. Could also be due to inorganics (SO_4", maybe PO_4")

- Used oil diluted 10:1 with pentane, centrifuged, supernatant decanted and precipitate washed with pentane.
    - Precipitate (as KBr disk)
        - + Calcium stearate
    - Oil after C_5 stripped
        - + Similar to oil before dilution but with much weaker soap peak.

- After calcium stearate had been precipitated with pentane, the oil still showed a peak at 1570 cm^{-1}. The oil was then shaken with dilute HCl, pentane added, the organic phase separated and dried, and the pentane stripped. The IR of the oil showed a strong increase in the 1715 cm organic acid band and the total elimination of the 1570 cm^{-1}, showing this latter band to be due to a soap. In addition, the 1100 cm^{-1} band also disappeared, again suggesting an inorganic ion as being responsible for this absorption. Some general weak absorption in the 1600-1700 cm^{-1} now shows (lost in the broad soap peak before) and this is probably due to some nitrogen-containing species.

No work was done on the aqueous extract.

JJE/bam
11/17/71

APPENDIX B

TABLE B-1

PILOT PLANT OPERATING PROCEDURE

- ### SLUDGE TRANSPORT AND STORAGE

The 900-gallon sludge tank, mounted on the rented flatbed truck, was filled at the sewage plant, returned to the pilot plant and allowed to settle as required to obtain the desired sludge concentration. After settling, the supernate was decanted off, the sludge recirculated with the process centrifugal pump to provide mixing, and the volume required for a run pumped up to the 300-gallon mixing tank. Mixing of the sludge tank contents by recirculation was only partially successful, due to channeling of supernate thru the settled solids; this presented no serious operating problems, but prevented the degree of presettling desired for some runs.

- ### FEED SLUDGE SAMPLING

After charging the prescribed run volume to the mixing tank, the contents were agitated for 1 minute and samples taken from the top and bottom (via dip samples and drainline, respectively) for % suspended solids. Sludge volume was determined from a previously prepared tank calibration.

- ### MIXING OF OIL AND SLUDGE FOR EXTRACTION

The oil to be used for the run was charged to the oil storage system and recirculated thru an external steam-heated heat exchanger until the oil temperature reached 240°F. The oil was then charged to the mix tank, thru the heat exchanger, at 250°F, with the quantity charged determined from the oil tank calibration curve.

The oil was charged to the sludge without agitation. After the oil has been added, the tank contents are thoroughly mixed by agitating for 10 seconds before starting the process (transfer and mixing) pump. The process pump is a Marlow open impeller centrifugal*, operating at 3460 RPM. Residence time in the pump was adjusted to ~1/3-1/2 seconds, comparable to laboratory operation, by adjusting the pump outlet throttle valve. The discharge from the pump was fed to the 500 gallon settling tank.

3-5# air pressure was put on the mixing tank during transfer to maintain feed rate and to blow the lines. After emptying the tank, an additional 5 gallons of oil were charged and added to the batch in the settler to clean the lines.

After completing the transfer, the oil and sludge lines were drained, and the drainings weighed for use in the material balance.

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* Same type as lab pump.

- SETTLING FOR SEPARATION OF WATER RAFFINATE

Before transfer the heat transfer fluid in the settler jacket (Dowtherm A) was adjusted to 180°F and maintained at this temperature during settling. Settling was continued for 21-26 hours for most runs, with the settled water phase drained off periodically. The quantity of raffinate was determined in a calibrated, agitated 40-gallon measuring tank and then dumped to the sewer. The water collected in the measuring tank was sampled (with agitation) to obtain a representative sample for analysis for each sampling period. Raffinate temperature was measured in the measurement tank.

At the completion of the run, as determined by levelling off of the raffinate volume-settling time curve, the oil sludge concentrate was drained out of the settler into drums for weighing and storage prior to shipment to Carver Greenfield. 25 grams of mercuric chloride ($HgCl_2$) were added as a perservative just prior to removal of the batch from the settler.

APPENDIX C

TABLE C-1

CARVER-GREENFIELD HEAT TRANSFER TEST RESULTS

1st Stage Concentration

Pilot Plant Run No.	Sludge Type	Oil Used in Esso Concentration	% NFS in Feed(1)	Oil/ Solids Wt. Ratio	Product Temp.°F	Oil/ Source Temp.°F	Vacuum Inches Hg	Circulation Rate GPM	Overall U Min.	Overall U Max.	Overall U Ave(2)	Exchanger Tube Fouling
1	Bergen County Activated	#4 Heating Oil	6.5	27/1	130	150-160	25	low (3)	22	49		None
2	"	#1 Varsol	5.1	8.3/1	N.A. - unstable operation with rapid Varsol distillation							
3	"	#4 Heating Oil	4.8	13.6/1	157-160	203-208	17	low	23	32		None
4	Wards Island Activated	#4 Heating Oil	4.3	46.8/1	160-170	180-190	17	low	6.8	37.5		None
5	"	#1 Varsol	4.0	9.5/1	120-170	148-152	25.5	low	41	60		None
6	Bergen County Primary & Activated	#4 Heating Oil	7.2	11.7/1	121-128	145-155	26	low	19	52		None
7	Trenton Trickle Filter	#4 Heating Oil	5.0	13/1	145-155	170-175	21.5	low	8	33		None
Hershey, Pa. primary & secondary (4)		Coray 37	5.0					Normal	93	122		None
Bergen County, N.J. Activated (4)		#2 Heating Oil	1.32	14/1	120	140	26.5	Normal	75	186		None

---

(1) % Nonfat solids = % feed solids corrected for solvent extractable fraction; % solids in water phase; Carver-Greenfield analysis.

(2) Design basis, representative of steady state conditions.

(3) Estimated at <2 gpm vs. normal 4.5-5.

(4) Samples directly from plant without prior treatment.

TABLE C-2

CARVER-GREENFIELD HEAT TRANSFER TEST RESULTS

Drying Stage Concentration

Pilot Plant Run No.	Sludge Type	Oil Used in Esso Concentration	% NFS in Feed(1)	Oil/ Solids Wt. Ratio	Product Temp.°F	Heat Source Temp.°F	Vacuum Inches Hg	Circulation Rate GPM	Overall U Min.	Overall U Max.	Overall U Ave(2)	Exchanger Tube Fouling
1	Bergen County Activated	#4 Heating Oil			240	280-290	19	4.5	54	58		None
2	"	#1 Varsol			N.A. →							
3	"	#4 Heating Oil			240-250	240	18	4.5	33	45		None
4	Wards Island Activated	#4 Heating Oil			229-240	260-280	19	4.5	75	97		None
5	"	#1 Varsol			235-244	262-271	19	4.5	50	110		None
6	Bergen County Primary & Activated	#4 Heating Oil			Not Run							
7	Trenton Trickle Filter	#4 Heating Oil			Not Run							
Hershey, Pa. priamry & secondary (4)		Coray 37			215	258	15	Normal	93	132		None
Bergen County, N.J. Activated (4)		#2 Heating Oil			215	260	15	Normal	93	130		None

---

(1) % Nonfat solids = % feed solids corrected for solvent extractable fraction; % solids in water phase; Carver-Greenfield analysis.

(2) Design basis, representative of steady state conditions.

(3) Estimated at <2 gpm vs. normal 4.5-5.

(4) Samples directly from plant without prior treatment.

TABLE C-3

DRY RECYCLE SOLIDS ADDED TO REDUCE VISCOSITY

- 1ST STAGE CONCENTRATION

PILOT PLANT RUN NO	% NFS IN FEED	OIL/SOLIDS WT. RATIO	PRODUCT TEMP. °F	HEAT SOURCE TEMP. °F	VACUUM INCHES Hg	CIRCULATION RATE, GPM	OVERALL U			EXCHANGER TUBE FOULING
							MIN.	MAX.	AVE.	
6A[1]	30	11.7/1	126	150	265	LOW	42	66		
7A	30	13/1	140	160	26	4.5	32	75	60	None

- DRYING STAGE CONCENTRATION

PILOT PLANT RUN NO	% NFS IN FEED	OIL/SOLIDS WT. RATIO	PRODUCT TEMP. °F	HEAT SOURCE TEMP. °F	VACUUM INCHES Hg	CIRCULATION RATE, GPM	OVERALL U MIN.	OVERALL U MAX.	OVERALL U AVE.	EXCHANGER TUBE FOULING
6A[1]			225-235	275	19	4.5	52	72	60	None
7A			230	260	19	4.5	132	210	160	

---

(1) Batch decomposed during 8 week storage before test; low U values associated with decomposition.

TABLE C-4

ANALYSIS OF DRIED SLUDGE SOLIDS FROM CENTRIFUGE

- "As is" (with oil) basis residual

PILOT PLANT RUN NO	TYPE SLUDGE	WEIGHT % OIL	WEIGHT % WATER
1	BERGEN COUNTY ACTIVATED	46.7	1.8
3	BERGEN COUNTY ACTIVATED	47.7	1.0
4	WARDS ISLAND ACTIVATED	40.7	1.8
5	WARDS ISLAND ACTIVATED	40.0	2.8
6	BERGEN COUNTY PRIMARY TACT	44.3	0.9

- Oil free basis

PILOT PLANT RUN NO	WEIGHT % ASH	WEIGHT % C	WEIGHT % H	WEIGHT % W	HEATING VALUE, BTU/# GROSS	HEATING VALUE, BTU/# NET(1)
1	46.7	25.8	3.8	4.6	4,856	4,509
3	47.4	25.2	3.8	4.5		
4	39.9	29.9	4.3	4.0	5,245	4,853
6	47.3	37.9	5.5	4.9		

(1) Gross BTU/# corrected for hydrogen according to procedure in ASTM D-2382.

TABLE C-5

ANALYSIS OF RECYCLE OIL

PILOT PLANT RUN NO.	WEIGHT %		VISCOSITY, SSU (100°F)(1)
	WATER	NON FAT SOLIDS	
1	<0.1	2.7	115.9
3	"	1.9	115.4
4	"	2.2	
5	"	2.4	
6	"	2.0	
7	"	2.2	

---

(1) Fresh oil viscosity = 73.1 SSU (100°F)

TABLE C-6

TEST DATA FROM
CARVER GREENFIELD CORPORATION

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - LITTLE FERRY

Date: 12/27/71 - 1/27/72

Run (customer)# 1 Carver-Greenfield # 641 Raw Feed: SECONDARY SEWERAGE SLUDGE

Oil Used For Drying: #4 FUEL OIL Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
37.3	2.36	64.0	6.3	27:1	27:1	NORMAL	FINE	NFS - 6.5%

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 92.5 %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vo
27:1	130	25	150-160	22	44		LOW	HIGH	NONE	-	9.7	SLIGHT NH_3	-	SOME

2ND STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 100 %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vol
	240	19	280-290	54	58		4.5	LOW	NONE	-	6.7	PETROL SLIGHT ROTTING	-	SOME

____ STAGE EQUILIBRIUM CONDITION

* Design Average

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vol

Remarks: 3 EFFECTS COULD NOT BE RUN AS VISCOSITY OF FLUID WAS TO HIGH. WENT DIRECTLY TO DRYING AFTER 1ST STAGE. 2/3 WATER REMOVED IN 1st STAGE.

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: Esso - Little Ferry

Run (customer)# 1 Carver-Greenfield # 641

Date: 10/27/81 - 1/27/82

DEHYDRATED SLURRY

Oil/Solids	% H_2O
20/1	TRACE

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 200 °F. Type BIRD G's 3000 Rate 2 GPM Pool Depth SHALLOW

% Solids	%H_2O	%Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
53.67	44.53	1.8										

RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
	2.7										

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure ____ Blowing Steam Rate ____

PRODUCT: Prod. out Temp.____°F. Production Rate ____ Heat Source Temp.____°F.

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks:

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: Essx - Little Ferry

Date: 1/28/72

Run (customer)# 3

Carver-Greenfield # 642

Raw Feed: SECONDARY ACTIVATED

Oil Used For Drying: VARSOL

Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H_2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
57.0	3.59	24.5	6.4	8.85 : 1	8.85 : 1	LOW	SMALL	NFS - 5.09

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 7.0 %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
				0	0									

____ STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
											6.8	PET. & SLIGHTLY [illegible]		

____ STAGE EQUILIBRIUM CONDITION

* Design Average

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo

Remarks: NOTE THAT UPON FEEDING MATERIAL, SUBSTANTIALLY ALL THE VARSOL CAME OFF CAUSING INABILITY TO KEEP SOLIDS IN SUSPENSION.

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: Esso - Little Ferry

Run (customer)# 3 Carver-Greenfield # 642

Date: 1/28/72

DEHYDRATED SLURRY

Oil/Solids	% H_2O

GRAVITY THICKENED "No Results"

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp._____ °F. Type _____ G's _____ Rate _____ Pool Depth_____

% Solids	%H_2O	%Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp._____°F. Pressure _____ Blowing Steam Rate _____

PRODUCT: Prod. out Temp._____°F. Production Rate _____ Heat Source Temp._____°F.

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks: _____

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSC - LITTLE FERRY

Date: 4/3 72 - 5 3 72

Run (customer)# 4

Carver-Greenfield # 503

Raw Feed: Secondary Activated

Oil Used For Drying: #4 FUEL OIL

Oil Present In Raw Feed: Sewerage Oil

RAW FEED:

%H_2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
51.5	8.91	39.7	6.3	13.6 : 1	13.6 : 1		SMALL	NFS - 4.83%

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 70 %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vc
13.6:1	157-160	17	203-208	23	32		LOW	HIGH	NONE		8.1	Slight NH3	-	Slight

2 STAGE EQUILIBRIUM CONDITION (DRYING)

NFS ____ %Design NFS 100 %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vc
	240-250	18	290	33			4.5	LOW	NONE		6.8	Petrol. & Slight [illegible]		Slight

____ STAGE EQUILIBRIUM CONDITION

* Design Average

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vc

Remarks: 2/3 Water removed in 1st STAGE.

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - LITTLE FERRY

Date: 1/31/72 - 2/3/72

Run (customer)# 2

Carver-Greenfield # 643

DEHYDRATED SLURRY

Oil/Solids	% H_2O
11.9:1	TRACE

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 200+ °F. Type Bird G's 3000 Rate 8GPM Pool Depth SHALLOW

% Solids	% H_2O	% Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
49.2	1.0	49.8										

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RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
	1.9										

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure ____ Blowing Steam Rate ____

Prod. out Temp.____°F. Production Rate ____ Heat Source Temp.____°F.

PRODUCT:

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks:

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSC - WARDS ISLAND

Date: 2-5-1/72

Run (customer)# 5 Carver-Greenfield # 342 Raw Feed: ACTIVATED SECONDARY

Oil Used For Drying: #4 HEATING OIL Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
32.1	1.4	66.5	6.0	34.8 : 1	46.8 : 1	NORMAL	SMALL	N.F.S. - 4.25%

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 7.75 %Actual %H2O - %Solids - %Oil -

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vo
	160-167	17	180-190	6.8	37.5		LOW	HIGH	NONE		9.0	SLIGHT NH3		

2nd STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 100 %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vol
	229-240	19	260-280	75	97		4.5	LOW	NONE		6.0	Aromatic & [illegible]		

____ STAGE EQUILIBRIUM CONDITION

* Design Average

NFS ____ %Design NFS ____ %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans. Min.	Overall Heat Trans. Max.	Overall Heat Trans. Avg.*	Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate Ph	Distillate Odor	Distillate COD	Distillate %Oil Vol

Remarks: ____

Page 1 of 2

"Test Data"

Customer: [illegible]

Date: 2/8-11/72

Run (customer)# 5

Carver-Greenfield # 644

DEHYDRATED SLURRY

Oil/Solids	% H_2O
19:1	TRACE

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 200+ °F. Type Bird G's 3000 Rate 8 GPM Pool Depth SHALLOW

% Solids	%H_2O	%Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
59.0	1.8	39.2										

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RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
	2.15										

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure ______ Blowing Steam Rate ________

PRODUCT: Prod. out Temp.____°F. Production Rate __________ Heat Source Temp.____°F.

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks:

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - WARDS ISLAND

Date: 8/15 - 28/72

Run (customer)# 6

Carver-Greenfield # 645

Raw Feed: SECONDARY ACTIVATED

Oil Used For Drying: VARSOL

Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
70	2.83	27.17	6.7	9.5:1	9.5:1	LOW	SMALL	NFS - 3.95%

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 7.35 %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
	120-140	25.5	148-152	41	60		LOW	HIGH	NONE		9.2	Slight NH3		38

2ND STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
	235-244	19	262-271	50	110		4.5	Normal	NONE		6.7	[illegible]		57

____ STAGE EQUILIBRIUM CONDITION

* **Design Average**

NFS ____ %Design NFS ____ %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo

Remarks: ____________________

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - [illegible] ISLAND

Date: 2/15-22 '72

Run (customer)# 6   Carver-Greenfield # 645

DEHYDRATED SLURRY

Oil/Solids	% H_2O
20:1	TRACE

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 200 °F. Type BIRD G's 3000 Rate 2 GPM Pool Depth SHALLOW

% Solids	% H_2O	% Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
57.2	2.8	39.9										

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RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
	2.35										

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure ____ Blowing Steam Rate ____

PRODUCT: Prod. out Temp.____°F. Production Rate ____ Heat Source Temp.____°F.

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks: ____

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO-LITTLE FERRY Date: 8/28-29/72

Run (customer)# 7 Carver-Greenfield # 546 Raw Feed: PRIMARY & SECONDARY

Oil Used For Drying: #4 HEATING OIL Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H_2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
52.3	3.74	43.96		11.7:1	11.7:1	NORMAL	SMALL	

1st STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp.°F.	Vac. "Hg.	Heat Source Temp.°F.	Overall Heat Trans.			Circ.Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
	121-128	26	125-155	19	82	O	LOW	HIGH	NONE	—				

____ STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp.°F.	Vac. "Hg.	Heat Source Temp.°F.	Overall Heat Trans.			Circ.Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo

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____ STAGE EQUILIBRIUM CONDITION

* Design Average

NFS ____ %Design NFS ____ %Actual %H_2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp.°F.	Vac. "Hg.	Heat Source Temp.°F.	Overall Heat Trans.			Circ.Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo

Remarks: SHUT DOWN TOO VISCOUS - SEE 7A

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - LITTLE FERRY

Run (customer)# 7A Carver-Greenfield # G66-A

Date: 4/24-26/72

Raw Feed: PRIMARY & SECONDARY

Oil Used For Drying: #4 FUEL OIL

Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H_2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
52.3	3.74	43.94		11.7:1	11.7:1	NORMAL	SMALL	

1st. STAGE EQUILIBRIUM CONDITION

NFS 30① %Design NFS 30① %Actual %H_2O — %Solids — %Oil —

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.②			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vc
	126	26.5	150	42	66		LOW	NORMAL	NONE		9.7	Petrol. NH_3	—	—

2ND STAGE EQUILIBRIUM CONDITION (DRYING)

NFS _____ %Design NFS 100 %Actual %H_2O_____ %Solids _____ %Oil _____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.②			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo
	225-235	19	275	52	72	60	4.5	NORMAL	NONE		5.45	Petrol & [illegible]	—	—

_____ STAGE EQUILIBRIUM CONDITION

NFS _____ %Design NFS _____ %Actual %H_2O_____ %Solids _____ %Oil _____

* Design Average

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vo

Remarks: ① BROUGHT TO 30% NFS TO SIMULATE THE CONDITIONS IN FIRST TWO STAGES (USING DRY SOLIDS FROM OTHER RUNS).

② LOW HEAT TRANSFER RATES ASSOCIATED WITH DETERIATED FEED.

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - LITTLE FERRY (PRIMARY & SECONDARY)

Date: 2/28-29/72

Run (customer) # 7A

Carver-Greenfield # 646

DEHYDRATED SLURRY

Oil/Solids	% H_2O
NOT RUN	

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 800+ °F. Type BIRD G's 3000 Rate 8GPM Pool Depth SHALLOW

% Solids	% H_2O	% Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
54.8	.9	44.3										

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RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%
	8.0										

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure _____ Blowing Steam Rate _______

Prod. out Temp.____°F. Production Rate _________ Heat Source Temp._____°F.

PRODUCT:

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks: ____________________

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO TRENTON TRICKLING FILTER Date: 2/29-3/16/72

Run (customer)= E Carver-Greenfield = G27 Raw Feed: SECONDARY TRICKLING FILTER

Oil Used For Drying: #2 FUEL OIL Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

H_2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
59.0	2.93	36.3	7.3	13:1	13:1	NORMAL	SMALL	

1ST STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS 6.6 %Actual %H_2O____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol.
	145-155	21.5	170-175	8	33		LOW	HIGH	NONE					

____ STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H_2O____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol.

____ STAGE EQUILIBRIUM CONDITION * Design Average

NFS ____ %Design NFS ____ %Actual %H_2O____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol.

Remarks: SHUT DOWN, TO VISCOUS - SEE 8A

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CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: ESSO - TRENTON - 1

Date: 2/29 - 3/7/72

Run (customer)# 8A Carver-Greenfield # 647 Raw Feed: SECONDARY TRICKLING FILTER

Oil Used For Drying: #4 FUEL OIL Oil Present In Raw Feed: SEWERAGE OIL

RAW FEED:

%H2O	%Solids	%Oil	Ph	Oil/NFS As Is	Oil/NFS As Feed	Viscosity	Particle Size	Remarks
59.0	2.93	38.3	7.3	13:1	13:1	NORMAL	SMALL	NFS - 4.75%

1st STAGE EQUILIBRIUM CONDITION

NFS 30① %Design NFS 30① %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol
	140	26	160	32	75	60	4.5	HIGH NORMAL	NONE					

2ND STAGE EQUILIBRIUM CONDITION (DRYING)

NFS ____ %Design NFS 100 %Actual %H2O ____ %Solids ____ %Oil ____

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol
	230	19	260	132	210		4.5	NORMAL	NONE					

____ STAGE EQUILIBRIUM CONDITION

NFS ____ %Design NFS ____ %Actual %H2O ____ %Solids ____ %Oil ____

* Design Average

Oil/NFS	Product Temp. °F.	Vac. "Hg.	Heat Source Temp. °F.	Overall Heat Trans.			Circ. Rate GPM	Visc.	Fouling	Act. Evap. Rate	Distillate			
				Min.	Max.	Avg.*					Ph	Odor	COD	%Oil Vol

Remarks: ① BROUGHT TO 30% NFS. TO SIMULATE FIRST 2 STAGES (USING DRY SOLIDS FROM PREVIOUS RUNS).

CARVER-GREENFIELD CORPORATION

"Test Data"

Customer: E-20 - TRENTON TRICKLING FILTER

Date: 2/29 - 3/16/72

Run (customer) # 8 Carver-Greenfield # 647

DEHYDRATED SLURRY

Oil/Solids	% H_2O

GRAVITY THICKENED

Time Period #1		Time Period #2		Time Period #3		Time Period #4		Temp. Maintained °F.
% Vol.	Time	% Vol.	Time	% Vol.	Time	% Vol.	Time	

CENTRIFUGE: Temp. 200 °F. Type BIRD G's 3000 Rate 2GPM Pool Depth SHALLOW

% Solids	% H_2O	%Oil	Fraction Of Oil									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

RECYCLE OIL:

% Solids		Fraction Of Oil									
Vol.	Wt.	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

PRESSING SOLIDS:

Temp °F.	Rate	% Oil	% H_2O

HYDROEXTRACTION: Prod. in Temp.____°F. Pressure ____ Blowing Steam Rate ____

Prod. out Temp.____°F. Production Rate ____ Heat Source Temp.____°F.

PRODUCT:

% Oil	% Sol.	% H_2O	Fraction Of Extractent									
			10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

Remarks: ____

Distillate	Fraction Of Extractent									
	10%	20%	30%	40%	50%	60%	70%	80%	90%	100%

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TABLE D-1

ENGINEERING, LEGAL AND ADMINISTRATIVE COST VS PLANT INVESTMENT

Total Construction Cost $MM (1)	Total Engineering Cost (2)		Legal, Fiscal and Administrative (3)		Combined	
	$M	%	$M	%	$M	% of Construction
.2	36	18	5.2	2.6	41.2	20.6
.5	70	14	9.2	1.8	79.2	15.9
1.0	120	12	14	1.4	134	13.4
2.0	200	10	22	1.1	222	11.1
3.0	270	9	27	0.9	297	9.9
4.0	340	8.5	31	0.8	371	9.3
5.0	410	8.2	35	0.7	445	8.9

---

(1) Construction cost = installed cost

(2) Ref. pg. 55 for complete plant

(3) Ref. pg. 57

TABLE D-2

CAPITAL COSTS FOR ESSO OIL CONCENTRATION PROCESS
SLUDGE PRETHICKENING - FINAL SETTLING AT 175°F

Raw Influent MGD	Secondary Waste Sludge MGD (1)	Secondary Solids Tons/day (2)	Waste Sludge Settling (3)	Thickened Sludge MGD	Thickener Design		Oil Sludge Mixer (5)		Concentration Factor (6a)	Oil-Sludge Heating (6)			Oil-Sludge Settler Cost-$MM (7)	Concentrate Surge Tank		Process Oil Storage - $MM		Process Pumps	
					Surface Area-ft² (4)	Installed Cost $MM (4a)	Cost-$MM	H.P.		Exchange Cost-$MM	Boiler Cost-$MM	Fuel Costs $/Ton Solids		Cost $MM (8)	Motor H.P.	Tank (9)	Oil (10)	Cost $MM	H.P.
12.5	.225	4.72	Fastest	.075	905	.046	.0046	2	8	.015	.012	3.75	.105	.025	2	.0106	.0025	.028	18.5
			Median		2,360	.068			6			4.50	"	.026	2				
			Slowest		3,780	.088			4			5.15	"	.028	3				
37.5	.675	14.16	Fastest	.225	2,710	.074	.0064	3	8	.029	.020	3.75	.200	.037	4	.0230	.0075	.042	41
			Median		7,080	.125			6			4.50	"	.042	5				
			Slowest		11,350	.173						5.15	"	.049	6				
250	4.5	94.5	Fastest	1.5	18,050	.248	.0085	6	8	.090	.065	3.75	.843	.164	10	.0700	.050	.110	193
			Median		47,300	.550			6			4.50	"	.177	11				
			Slowest		76,600	.840			4			5.15	"	.207	12				

---

1) 1.8% of influent.
2) 0.5-.6% suspended solids in waste sludge.
3) Sludge thickened to 1.5%.
4) Calc. from batch setting data and Kynch Method.
4a) Cost data in Appendix D-4.
5) In-line type, high shear.
6a) Equivalent to 5 hrs at 105°F and 15 hrs at 175°F.
6) Heating from 105°F to 175°F before 3rd stage settling.
7) 3 stage settling; see Appendix D-5 for details.
8) 24 hour holding capacity of oil sludge concentrate for evaporation; includes agitator; Carver-Greenfiedl design requirement.
9) Requirement for 12 hour reserve inventory.
10) Oil cost at $0.11/gal; 24 hr. process inventory + 12 hr. reserve.

TABLE D-3

COST ESTIMATE FOR ESSO OIL CONCENTRATION PROCESS
SLUDGE PRETHICKENING - FINAL SETTLING AT 175°F

Plant Size		Process Operation		Investment - $MM				Costs - $/Ton Feed Solids							
Raw Sewage Influent MGD	Secondary Solids Tons/Day	Sludge Thickening Rate	Final Solids Content	Installed	TIE (1)	$MM/Ton	H.P. for Motors	Capital	Maintanence	Insurance	Power	Fuel	Labor + Overhead	Total TC Loss	Total
12.5	4.72	Fastest	Max.	.250	.322	.0682	22	13.25	9.35	1.87	.67	3.75	14.86	4.37	48.12
		Median	Median	.273	.351	.0744		14.45	10.20	2.05		4.50	"		51.10
		Slowest	Min.	.295	.380	.0804		15.60	11.00	2.21		5.15	"	"	53.86
37.5	14.12	Fastest	Max.	.439	.552	.0389	49	7.55	5.34	1.07	.50	3.75	4.84	"	27.42
		Median	Median	.495	.623	.0440		8.55	6.03	1.21		4.50	"		30.00
		Slowest	Slowest	.550	.694	.0481		9.35	6.60	1.33		5.15	"	"	32.14
250	94.5	Fastest	Max.	1.65	2.00	.0212	210	4.10	2.90	.59	.32	3.75	1.49	"	17.5
		Median	Median	1.96	2.38	.0252		4.90	3.45	.63		4.50	"		19.7
		Slowest	Slowest	2.28	2.76	.0292		5.68	4.00	.821		5.15	"	"	21.8

(1) Summation of installed equipment cost, - 10% contingency, engineering + legal + adminstrative (see Table D for factors).
(2) Based on Chicago influent BOD charge and recycle factors as detailed in Attachment
(3) Based on 25% total TC recycle; includes both settling and evaporation steps.
(4) 1/2 man/shift for 4.72 and 14.12 T/D plants, 1 man/shift for 94.5 T/D plant.

TABLE D-4

COST OF SLUDGE THICKENERS AND OIL-SLUDGE SETTLERS

Surface Area 1000 ft^2	Construction Costs - \$M (1) Jan 1971 (2)	Construction Costs - \$M (1) Spring 1972 (3)
1	41	43
2	55	62
5	91	102
10	140	157
20	240	269
40	430	482
60	600	674
80	760	853
100	920	1032

---

(1) Construction cost = installed cost

(2) Ref. 37 pg. 37

(3) Corrected for 12% inflation factor to March 1972, based on Sewage Treatment Plant Construction Cost Index, Ref. 41.

TABLE D-5

INSTALLED COST OF OIL SLUDGE SETTLERS

Plant Size MGD Thickened Sludge	1st Stage Drum Cost $MM (1)	2nd Stage				3rd Stage				Combined Settler Cost Installed-$MM
		Area 1000 ft^2	Installed Cost-$MM	Roof Cost(2)	Total Installed	Area 1000 ft^2	Installed Cost	Roof Cost	Total Installed	
.075	.0052	.782	.045	.005	.050	1.35	.054	.006	.060	.105
.225	.0134	2.34	.068	.010	.078	4.05	.091	.017	.108	.200
1.5	.0450	16.05	.223	.074	.297	27.70	.351	.150	.501	.843

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(1) Cost data from ref. - for drum settler, corrected for horizontal plates on basis of cost data from Esso Engineering

(2) Cost data from Esso Engineering.

TABLE D-6

COST ESTIMATE FOR CARVER GREENFIELD PROCESS

Plant Size		Type Operation shifts/day	Operators per shift (2)	% Solids in Concentrate	Evaporator Operation		Total Installed & Erected Cost - $MM			Costs - $/Ton Feed Solids						
Raw Influent MGD	Sludge Solids Tons/Day (1)				Effects	U Value (2a)	Installed (3)	TIE	$MM/Ton/Day	Capital (4)	Maintenance (5)	Insurance (6)	Total Labor (7)(8)	Power (9)	Fuel (10)	Total
.225	4.72	1	1	4.5	3	60(11)	.870	.883	.187	36.4	25.6	5.15	9.87	3.31	4.45	84.78
	"			9.0	"	"	.630	.604	.136	26.5	18.6	3.74		2.02	(2.24)	58.49
.225	4.72	3	1	4.5	3	60	.605	.615	.130	25.3	17.8	3.58	29.70	3.32	4.45	84.15
				6.0	"	"	.575	.585	.125	24.3	17.1	3.41		2.87	2.20	79.58
				9.0	"	"	.550	.560	.119	23.2	16.3	3.28		2.84	(2.24)	73.08
				4.5	"	120	.572	.580	.123	23.9	16.9	3.38		2.50	4.45	80.83
				9.0	"	120	.539	550	.117	22.8	16.0	3.22		2.84	(2.24)	72.42
.675	14.2	3	1	4.5	"	"	.870	.883	.0622	12.1	8.52	1.72	9.87	3.31	4.45	39.97
				6.0	"	"	.767	.780	.0552	10.7	7.56	1.52		2.87	2.20	33.72
				9.0	"	"	.630	.640	.0452	8.76	6.18	1.24		2.02	(2.24)	25.83
				4.5	"	120(12)	.823	.835	.0590	11.5	8.08	1.63		2.30	4.45	38.03
				9.0	"	120	.590	.600	.0425	8.26	5.82	1.17		2.02	(2.24)	24.90
				4.5	4	"	.810	.822	.0581	11.3	7.96	1.61		2.54	(.16)	33.12
				6.0	4	"	.705	.716	.0506	9.85	6.95	1.40		2.30	(3.04)	27.32
4.50	94.5	3	3	4.5	3	60	3.80	3.83	.0406	7.90	5.56	1.12	4.47	3.30	4.45	26.80
			2	6.0	"	"	3.18	3.21	.0340	6.61	4.66	0.94		2.87	2.20	20.25
				9.0	"	"	2.35	2.38	.0252	4.90	3.45	0.70		2.02	(2.24)	11.80
				4.5	3	120	2.80	2.83	.0300	5.48	4.12	0.83	2.97	2.50	4.45	20.35
				9.0	3	120	1.85	1.87	.0198	3.87	2.73	0.55		2.02	(2.24)	9.90
				4.5	4	120	2.58	2.61	.0280	5.45	3.85	0.77		2.25	(.16)	15.13
				6.0	4	120	2.08	2.10	.0222	4.32	4.32	0.61		2.20	(3.04)	10.10

(1-10) See attachment for description of footnotes.

FOOTNOTES FOR TABLE D-6

1. Based on waste sludge = 1.8% of influent volume, with suspended solids content of 0.50%.

2. Process operators specified by Carver Greenfield.

2a. Overall heat transfer coefficient, BTU/hr/ft^2/°F.

3. Quotation from Carver Greenfield; complete erected cost, including boiler; includes 10% contingency.

4. Based on interest and amortization for 25 years, 5% interest rate on bonds.

5. 5% of total investment.

6. 2% of total investment.

7. Operating labor cost = \$3.90/hr direct labor cost + 15% for indirects = \$4.50/hr.

8. Taken as 30% of operating labor.

9. Based on electric power cost of \$.010/kwh, and usage at 90% of installed H.P.

10. Based on fuel oil cost of \$.016/#. Where excess energy produced from incineration, credited at equivalent fuel value; excess heat denoted by ( ).

11. Value obtained from Carver Greenfield heat transfer studies on Esso oil sludge concentrates.

12. Value assumed for low temperature settling, with no viscosity limitation during evaporation.

TABLE D-7

FUEL AND POWER REQUIREMENTS FOR CARVER GREENFIELD EVAPORATION PROCESS (1)

Plant Size Tons/Day Sludge	Weight % Water in Feed	Number of Effects	U Value	Net Heating Value of Sludge Solids	Total Horsepower Installed	Fuel Requirements #/Day (2)
4.72	4.5	3	60	4	110	1580
	9.0	"	"	"	94	(495)
14.17	4.5	"	"	"	31.8	4750
	9.0	"	"	"	20.0	(1490)
	4.5	4	120		220.8	552
	6.0	"	"	"	13.61	(2040)
94.5	4.5	3	60	4	25.2	32,100
	9.0	"	"	"	20.1	(7850)
	4.5	4	120	"	148.7	3860

(1) Carver-Greenfield data.
(2) ( ) denotes excess energy expressed as fuel equivalent.

TABLE D-8

PRELIMINARY HEAT AND MATERIAL BALANCES FOR THE CARVER GREENFIELD DEHYDRATION PROCESS

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

DATE May 8, 1972

MATERIAL TO BE DRIDED

PROPOSAL NO. 072-0077-1

REF. NO.

Rate of Feed 10,625 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	90.5	9750	4.6	45
Solids	9.0	875	89.0	875
Oil in Feed	.5	118	6.4	60
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect Steam: 4.07 x 106 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 199-1/2

Cooling Water: Sludge Thickening Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn as fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 4.97 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 58 overage Lbs. ~~Gals~~/Hr.

Total Fuel if Solids are Recovered -- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 25 Ft., Wd. 30 Ft., Ht. 75 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled) $450,000 Max. boiler cost
$410,000 Min. boiler cost
Est. Install. Cost $200,000
" Stack Emission $10,000

D-11

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-2

REF. NO.

Rate of Feed 19,421 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	4-1/2	18,646	4.2	49
Solids	95-1/2	875	75	875
Oil in Feed			20.8	240
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect Steam: 7.78 x 106 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 318

Cooling Water: Sludge Thickening Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn for fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 4.97 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 190 Lbs. XXXXX/Hr.

Total Fuel if Solids are Recovered --- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 35 Ft., Wd. 30 Ft., Ht. 70 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled)
$625,000 Max. boiler cost
$565,000 Min. boiler cost
Est. Install. Cost $275,000
Est. Emission Cost $10,000

D-12

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-3

REF. NO.

Rate of Feed 130,860 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	95-1/2	125,000	3.3	270
Solids	4-1/2	5,860	72.5	5860
Oil in Feed			24.2	1945
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect Steam: 52 x 106 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 2208

Cooling Water: Sludge Thickening Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn for fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 323 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 1340 Lbs. xxxxs/Hr.

Total Fuel if Solids are Recovered Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 90 Ft., Wd. 90 Ft., Ht. 75Ft.

Manpower Requirements 96 Hrs./Day

Estimated Sales Price (Uninstalled)
$2,800,000 Max. Boiler cost
$2,400,000 Min. Boiler cost
Est. Install. $1,200,000
Est. Emission $40,000

D-13

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-4

REF. NO.

Rate of Feed 68,350 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	91	62,500	4.1	270
Solids	9	5,850	89.0	5850
Oil in Feed			7.9	138
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect Steam: 28.15 x 106 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 1361

Cooling Water: Sludge Thickening Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn for fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 33 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 327 Lbs. Excess ~~Gals~~/Hr.

Total Fuel if Solids are Recovered -- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 30 Ft., Wd. 90 Ft., Ht. 75 Ft.

Manpower Requirements 48 Hrs./Day

Estimated Sales Price (Uninstalled)
$1 00,000 Max. Boiler Cost
$1,500,000 Min. " "
Est. Install. $700,000
Est. Emission $40,000

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DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-5

REF. NO.

Rate of Feed 6493 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	94.5	6200	3.8	15
Solids	4.5	293	74.5	293
Oil in Feed	.5	39	21.7	83
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect
Steam: 2.63 x 103 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 110

Cooling Water: Heating up of sludge Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn as fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 2.06 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 44 Lbs. ~~Gals~~/Hr.

Total Fuel if Solids are Recovered -- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 20 Ft., Wd. 20 Ft., Ht. 75 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled) $615,000 Max. boiler cost
$595,000 Min. boiler cost
Est. Install. $195,000
Est. Emission $10,000

DESIGN CRITERIA FOR
THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-6

REF. NO.

Rate of Feed ____ Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	90.5	2950	4.3	15
Solids	9.0	293	86.7	293
Oil in Feed	.5	39	9.0	39
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

3 Effect
Steam: 1.24 x 106 BTU/Hr. ____ Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 94

Cooling Water: Heating up of sludge Gals./Hr. ____ Gals./Min.

Ultimate Use of Solids Burn as fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 2.06 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value -- Gals/Hr.

Total Fuel if Solids are Recovered -- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 20 Ft., Wd. 20 Ft., Ht. 20 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled) $560,000 Max. boiler cost
$540,000 Min. boiler cost
Est. Install. $175,000
Est. Emission $10,000

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-7

REF. NO.

Rate of Feed 21,540 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	95	20,400	4.4	51
Solids	4.5	960	83.5	960
Oil in Feed	.5	117	21.1	140
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

4 Effect
Steam: 6.475 x 106 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 252

Cooling Water: Sludge Thickening Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn as fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 6.15 x 106 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 23 Lbs. ~~Gals~~/Hr.

Total Fuel if Solids are Recovered ---- Gals/Hr.

General Material of Construction of Equipment Carbon Steel

Approx. Bldg. Space: Lg. 44 Ft., Wd. 22 Ft., Ht. 75 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled)
$590,000 Max. boiler cost
$530,000 Min. boiler cost
Est. Install. $250,000
Est. Emission $10,000

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DESIGN CRITERIA FOR THE CARVER GREENFIELD PROCESS

CUSTOMER ESSO-EPH

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-8

REF. NO.

Rate of Feed ______ Lbs./Hr. Est. Hrs./Day Operation ______

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	93.5	15,090	4.5	51
Solids	6.0	960	85.2	960
Oil in Feed	0.5	117	10.3	117
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

Steam: 4.91×10^6 BTU/Hr. ______ Lbs./Hr. @ ______ PSIG

4 Effect

Total Connected Horsepower 201

Cooling Water: Sludge Thickening Gals./Hr. ______ Gals./Min.

Ultimate Use of Solids Burn As Fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 6.15×10^6

Additional Fuel Required if Solids are for Fuel Value - Gals./Hr.

Total Fuel if Solids are Recovered - Gals/Hr.

General Material of Construction of Equipment Carbon Steel (C.S.)

Approx. Bldg. Space: Lg. 42 Ft., Wd. 22 Ft., Ht. 75 Ft.

Manpower Requirements 24 Hrs./Day

Estimated Sales Price (Uninstalled)	$520,000	(Max Boiler Cost)
	$430,000	(Min Boiler Cost)
Estimated Cost Of Install.	$220,000	
" " " Emmission	$ 10,000	

D-18

DESIGN CRITERIA FOR THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO-EPH

DATE May 8, 1972

MATERIAL TO BE DRIED

PROPOSAL NO. 072-0077-9

REF. NO.

Rate of Feed 139,800 Lbs./Hr. Est. Hrs./Day Operation 24

	ANALYSIS - FEED		ANALYSIS - DRIED PRODUCT	
	Percent	Lbs./Hr.	Percent	Lbs./Hr.
Water	95.0	132,500	4.3	323
Solids	4.5	6,250	83.0	6,250
Oil in Feed	0.5	758	12.7	923
Recycle Oil Rate				

Oil used for Fluidizing

Energy Requirements

4Eff.Steam: 42.81×10^6 BTU/Hr. Lbs./Hr. @ 100 PSIG

Total Connected Horsepower 1,487

Cooling Water: Heating Of Sludge Gals./Hr. Gals./Min.

Ultimate Use of Solids Burn as fuel

Total Fuel Value of Solids @ 80% Boiler Efficiency 40.4×10^6 BTU/Hr.

Additional Fuel Required if Solids are for Fuel Value 165 lb. ~~Gals~~/Hr.

Total Fuel if Solids are Recovered - Gals/Hr.

General Material of Construction of Equipment Carbon Steel (C.S.)

Approx. Bldg. Space: Lg. 80 Ft., Wd. 60 Ft., Ht. 75 Ft.

Manpower Requirements 48 Hrs./Day

Estimated Sales Price (Uninstalled) $2,100,000
$1,650,000

Estim. Install. Cost $610,000

Estim. Stack Emmiss. $ 40,000

TABLE D-9

PROJECTED COST SAVINGS FOR LOW TEMPERATURE SETTLING (1)

Plant Size Tons Sludge/Day	Concentration Factor	Fuel for Sludge Heating (3)	Superheat Equivalent (4)	TOC Recycle Charge (5)	Investment for Sludge Heating (6)	Total Esso Process (7)
4.72	Max.	-3.75	+.56	-2.53	-2.7	-8.4
	Min.	-5.15	+.90	"	"	-9.5
14.16	Max.	-3.75	+.56	"	-1.6	-7.3
	Min.	-5.15	+.90	"	"	-8.4
94.5	Max.	-3.75	+.56	"	- .72	-6.4
	Min.	-5.15	+.90	"	"	-7.5

(1) For Esso oil concentration process for secondary sludge only.
(2) - = cost decrease for 105°F settling.
    + = cost increase for 105°F settling.
(3) To heat oil sludge concentrate to 3rd stage settler from 105°F to 175°F.
(4) Fuel value of 3rd stage oil sludge concentrate at 175°F flashing to 105°F in 3rd stage evaporator.
(5) Charge for TC recycle or basis of BOD equivalent.
(6) Capital charges for boiler and heat exchanger,required to heat 2nd stage oil sludge concentrate from 105°F to 175°F.
(7) No change projected for operating manpower.

TABLE D-10

COST SAVINGS FOR 50% REDUCTION IN AREA OF OIL SLUDGE CONCENTRATORS (SETTLERS)

	2nd Stage				3rd Stage								
		Installed Cost $MM				Installed Cost							
Plant Size MGD Thickened Sludge	Reduced Area 1000 ft^2	Settler	Roof	Total	Reduced Area 1000 ft^2	Settler	Roof	Total	Total Installed Settler Cost	Installed Cost Savings Over Base Case	Factor for TIE (1)	Cost Savings TIE Basis	Cost Savings $/Ton (2)
.075	.39	.036	.004	.040	.675	.044	.004	.048	.088	.017	1.28	.022	1.6
.225	1.17	.052	.006	.058	2.02	.063	.008	.071	.129	.071	1.155	.090	2.3
1.5	8.03	.138	.035	.173	13.85	.200	.062	.262	.435	.408	1.11	.50	1.9

(1) Using engineering, legal factors for "Low" Esso Investment for base case.
(2) Using all investment based cost factors; for capital, maintanence, insurance.

TABLE D-11

CARVER-GREENFIELD CORPORATION
9 GREAT MEADOW LANE
HANOVER, NEW JERSEY 07936
—
(201) 887-2182

PRELIMINARY ESTIMATE
FOR
THE CARVER-GREENFIELD DEHYDRATION PROCESS

9% SOLIDS IN FEED
4.72 TONS/DAY SLUDGE SOLIDS

CUSTOMER: ESSO - EPA

PROPOSAL NO. 072-0077-1

DATE July 11, 1972

SF-8
8/71

GENERAL SPECIFICATIONS
FOR
THE CARVER-GREENFIELD PROCESS

CUSTOMER ESSO - EPA

PROPOSAL NO. 072-0077-1
DATE 7/11/72

ITEM	NO. SUPPLIED	MAT'L OF CONST.	HP TOTAL	REMARKS	
1. Raw Feed Holding Tank	--			Heated	
2. Raw Feed Tank Agitator	--				
3. Raw Feed Pump	--				
4. Fluidizing Tank	1	Carbon Steel	--	Heated	
5. Fluidizing Tank Agitator	1	"	1/2		
6. Fluidizing Pump	1	Cast Iron	3		
7. Fine Grinder	1	C.S/ S.S.	10		
8. Feed Tank	1	Carbon Steel	--	Heated	
9. Feed Tank Agitator	1	"	1/2		
10. Evaporator Feed Pump	1	Cast Iron	3		
11. Evap.-Concentrating Style	--				
12. Evaporator-Drying Style	1	Carbon Steel	--		
13. Circulation Pump(s)	3	Cast Iron	2@20 1@15		
14. Vapor Condenser (Barometric)	1	Carbon Steel			
15. Vapor Condenser (Surface)	--				
16. Vacuum Pump	--				
17. Vapor Line Preheater	1	Cast Iron	7-1/2		
18. Condensate Pump(s)	--				
19. Transfer Pump(s)	2	Cast Iron	2@1		
20. Product Pump	1	"	3	Vari-speed	
21. Centrifuge (Continuous)	1	Carbon Steel	25	12x30	
22. Centrifuge (Batch)	--			Auto. Man.	
23. Solids Bin or Tank	1	Carbon Steel	3		

GENERAL SPEC'S.-(con't.)

CUSTOMER ESSO- EPA

PROPOSAL NO. 072-0077-1

DATE 7/11/72

ITEM	NO. SUPPLIED	MAT'L OF CONST.	HP TOTAL	REMARKS	
24. Recycle Oil Tank	1	Carbon Steel	--	Heated	
25. Recycle Oil Pump	1	Cast Iron	1	Vari-speed	
26. Chutes	2	Carbon Steel	--		
27. Cooling Water Tower	--			W.B. °F D.B. °F	
28. Cooling Water Pump	1	Cast Iron	20		
29. Condensate System (18)	--				
30. Scalping Oil Tank	--				
31. Coalescer	--				
32. Holding Tank	--			Heated	
33. Holding Tank Agitator	--				
34. Bulk Oil Holding Tank	1	Carbon Steel	--		
35. Oil Pump	1	Cast Iron	1/2		
36. Operating Panel	1	Carbon Steel	--		
37. Controls	1 set	C.S/S.S.			
38. Piping & Fittings	1 set	Carbon Steel			
39. Motor Control Center	1				
40. Condensate Sump Pump	--				
41. Scalping Tank Discharge Pump	--				
42. Repulping Tank	--			Heated	
43. Repulping Tank Agitator	--				
44. Packaged Boiler **	--				
45. Boiler (Solids Handling)	1	Carbon Steel	30		
46. Multi-Compartment Hot Well	--				
47. Coarse Grinder	--				

GENERAL SPEC'S.-(con't.)

CUSTOMER ESSO - EPA

PROPOSAL NO. 072-0077-1
DATE 7/11/72

ITEM	NO. Supplied	MAT'L OF CONST.	HP TOTAL	REMARKS	
48. Bar Screen	--				
49. Desolventizer	1	Carbon Steel	10		
50. Screw Conveyor	1	"	3		
51. Crystallizer Tank	--			Heated	
52. Crystallizer Tank Agitator	--				
53. Fine Screen	--				
54. Cyclone Separator Liquid-Solid	--				
55. Boiler Feed Water System	1	Standard	3		
56. Filter	--				
57. PH Controller	--				
58. Silo Holding Bin	1	Carbon Steel	5		
59. Boiler Water Softening	1	Standard	2		
60. Deaerator & Feed Water Pumps	1 set	"	2@5		
61. Ash System	1	"	2		
62. Boiler Stack Emission Devices	1*		2		
63. Process Water Treatment					
*Not included in basic quotation					

NOTE:** If solids are to be recovered and heat source to be supplied by C-G.

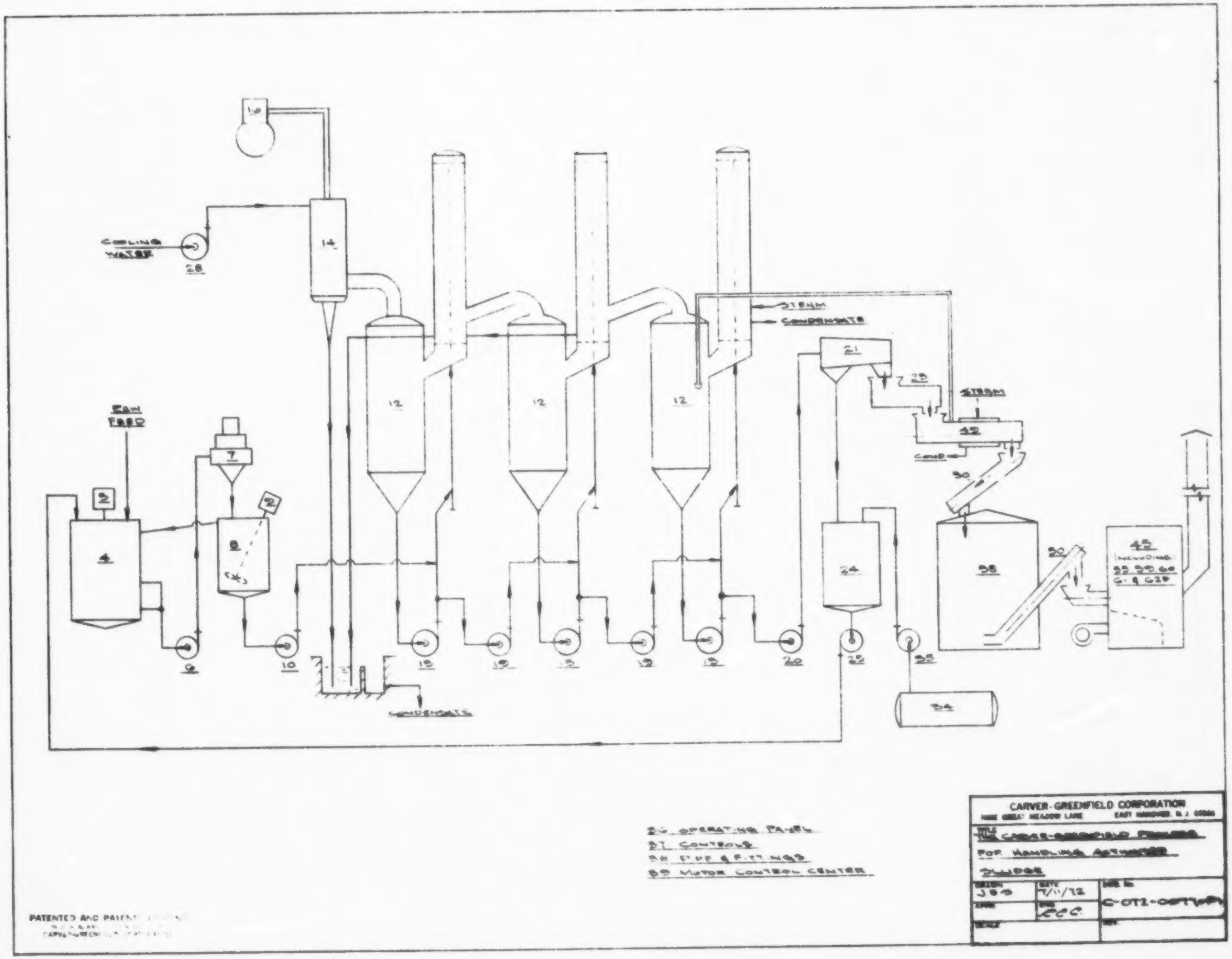
Process design is considered commercially acceptable. The addition of other features as required by various state, local, and governmental regulations, or by insurance companies will be considered beyond the scope of this proposal.

Carver-Greenfield Corporations proposal includes supplying the following:

1. Full process engineering services for equipment supplied by Carver-Greenfield Corporation, except structural or architectural engineering.
2. Engineering service liaison during construction of plant and start-up assistance.
3. Full set of operating instructions.
4. Guarantees of performance as ascertained from our own pilot plant study.

Normal Supplier Of Major Equipment

Major Equipment	Type	Supplier
Pumps	Centrifugal	Worthington
Evaporator	Falling Film	Mojonnier
Cooling Tower	Packaged	Marley
Centrifuge	Horizontal Bowl	Bird Machine
"	Basket	Fletcher, or Sharples
Motor Control Center	Modular	Allen-Bradley
Control Valves	Pneumatic	Masoneilan
Process Valves	Plug	ACF
Utility Valves	Globe & Gate	Crane
Piping	Welded	Tube Turn
Instrument Panel	Hoffman Box	C-G Subcontract
Pumps	Gear	Blackmer
Tanks	Vertical	C-G Subcontract
Agitators	Propeller	Mixing Equipment
Boiler-Furnace		Babcock & Wilcox, E. Keeler


COOLING WATER
RAW FEED
STEAM
CONDENSATE
STEAM
COND
CONDENSATE
2
4
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7
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24
25
28
30
32
45 (INCLUDING 55 56 60 61 & 62)
54
56 OPERATING PANEL
57 CONTROLS
58 PIPE & FITTINGS
59 MOTOR CONTROL CENTER
CARVER-GREENFIELD CORPORATION
NINE GREAT MEADOW LANE EAST HANOVER, N.J. 07936
TITLE THE CARVER-GREENFIELD PROCESS
FOR HANDLING ACTIVATED
SLUDGE
JGS
7/11/72
RCC
C-072-OPT
PATENTED AND PATENTS PENDING